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April 1987

Economic Feasibility Study of an Enzyme-Based Ethanol Plant

A Subcontract Report

Stone & Webster Engineering Corp.
Boston, MA

Prepared under Subcontract No. ZX-3-03097-1

SERI

Solar Energy Research Institute
A Division of Midwest Research Institute
1617 Cole Boulevard
Golden, Colorado 80401

Operated for the
U.S. Department of Energy
under Contract No. DE-AC02-83CH10093

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EXECUTIVE SUMMARY

1.1 INTRODUCTION

Stone & Webster Engineering Corporation (SWEC) has prepared an Economic Feasibility Study of an enzyme-based ethanol plant as requested by the Solar Energy Research Institute (SERI). The study's objectives are to determine the current economic status of the conversion of lignocellulose to ethanol via enzymatic hydrolysis and to provide recommendations for further research and development. The results of this study include an integrated process design, a capital cost estimate, an investment analysis (including an analysis of alternative designs), and recommendations for the direction of future research and development.

The site for the enzyme-based ethanol plant (EEP) is on the island of Hawaii, near the city of Hilo. The full scale plant will be capable of producing 15,000,000 gallons per year of fuel-grade ethanol from eucalyptus wood chips.

The technical information utilized in this study was obtained from SERI, published literature, The Hawaii Natural Energy Institute (HNEI), process and equipment vendors, and SWEC in-house data.

1.2 PROJECT SUMMARY

Ethanol is a high-value liquid fuel which can be derived from renewable, nonpetroleum feedstocks. Ethanol, as a beverage, has been produced for centuries from starch and sugar-bearing fruits, vegetables, and grain products. Unfortunately, these feedstocks are primarily used for both human and animal consumption, and are considered both economically and politically unattractive as feedstocks for fuel ethanol plants on a large scale ($\sim 10 \times 10^9$ gal/yr) application, such as that contemplated for the U.S. automotive fuel market. An alternative biomass feedstock which is not part of the basic food chain is wood. The technical and economic feasibility of commercially producing ethanol from wood via enzyme hydrolysis is assessed and reported in this study.

A specific site (near Hilo, on the island of Hawaii) was chosen to assess the technical and economic viability of producing ethanol from cultivated eucalyptus tree farms. Of the 2.5 million acres of land on the island of Hawaii, 569,000 acres are classified as forest land for commercial use. Studies performed by HNEI report that the eucalyptus species exhibit high growth rates that translate into a mean annual incremental yield of 10 bone-dry tons (BDT) per acre per year. This high feedstock growth rate is capable of easily supporting the requirements of the enzyme hydrolysis plant and is not a limiting factor in plant size determination. The plant size of 15 million gallons per year ethanol was selected as a maximum feasible size because of the market demand for fuel ethanol in the state of Hawaii. At the growth rate of 10 BDT/acre-yr, the enzyme-based ethanol plant (15 million gal/yr with the required 463,500 tons/yr of feedstock) would require 23,000 acres of Hawaii's forest land.

The eucalyptus wood contains three main fractions; cellulose (hexose), hemicellulose (pentose), and lignin. The base-case design uses only the hexose fraction of the wood to produce ethanol. The other wood components have the potential of producing significant revenue for the process. The pentose sugars produced in the enzyme hydrolysis process have a potential value as animal feed or could be converted to the by-product furfural. Lignin is also another potentially valuable by-product that could be sold to increase plant revenue. Unfortunately, the market for these by-products is undefined and further investigation would be required to determine their potential marketability and/or their export value. An alternative use for the pentose sugars would be to convert them directly to alcohol via fermentation. Although this would be considered to be economically favorable, pentose (C_5) fermentation is not yet considered to be a commercial technology. In view of the nonexistent by-product markets in Hawaii and commercially unproven technology, the base-case design utilizes the pentose sugars in an anaerobic digester to produce methane-rich gas. This methane-rich gas, along with extracted lignin, is used as boiler fuel to produce steam required by the process.

1.2.1 Design Criteria and Basis of Design

The design of the base-case EEP was determined by utilizing the following general guidelines:

1. Maximize the production of ethanol while minimizing by-product production.

2. Use commercially available equipment where applicable.

3. Assume technically undeveloped equipment (when used) can be scaled to commercial size.

4. Select base-case equipment, based on good engineering practice, which is conservative in actual design but optimistic in expected performance.

5. Maximize process heat integration and water recycle.

The base-case plant was designed and costed to determine a realistic current price for wood-derived ethanol. Although the individual processing sections are largely based on laboratory or bench-scale data, the enzyme hydrolysis base-case design is an integrated process scaled to a 15 million gal/yr size. The basis of design is summarized in Table 1.2-1.

The feedstock handling system receives chipped wood from the eucalyptus tree farm, removes metal and oversized material, contains 14 days of storage, and supplies a sized product to the process. The processed chips are conveyed to the pretreatment section where the wood chips are impregnated with a 0.2 wt percent concentration of sulfuric acid. This pretreatment step enhances the desirable effects of the steam explosion treatment (higher glucose yields), while minimizing the severity of the steam explosion conditions. The acid-soaked wood chips are fed into the steam explosion guns where high pressure steam is injected to raise the temperature and pressure to 464°F and 470 psig, respectively. The wood chips are cooked for 5 seconds at temperature and exploded into a medium pressure (MP) flash vessel. The steam explosion treatment enhances the accessibility of the wood to enzyme hydrolysis by disrupting the cellulose microfibrils, detaching the lignin, and increasing the surface area of the cellulose. The energy of steam explosion is recovered as 60 psig steam in the MP flash vessel for process use. The exploded pulp is cooled in a vacuum flash and then washed in the counter-current water/alkali wash to remove soluble hemicellulose (pentose) sugars and extract lignin with sodium hydroxide. The water wash is essential to remove the water soluble degradation products that are inhibitors of yeast fermentation and the production and activity of the enzyme complex. Lignin is removed prior to hydrolysis because of its potential interference with the enzyme recovery and hydrolysis performance. The recovered lignin is combusted as boiler fuel for the production of process steam. Eight percent of the washed cellulose stream is conveyed to the enzyme production section where the total hydrolysis makeup enzyme requirement is produced. The system design utilizes the Rut-C-30 strain of Trichoderma Reesei and is a batch fed enzyme fermentation modeled after experimental data developed at the University of California, in Berkeley. Fermentable sugars are formed by enzyme attack on the remaining washed cellulose stream in the hydrolysis section. The glucose concentration in hydrolysis was chosen as 7 weight percent with an anticipated yield of 84 percent of cellulose conversion within 48 hours. This information is based on data generated from a steam-exploded corn stover at the University of California, Berkeley, since information on steam-exploded eucalyptus is not readily available. Enzyme recovery is also included as part of the base-case design, which assumes 50-percent recovery of enzyme filter paper activity. The hydrolysate stream is concentrated in a five-stage, multi-effect evaporator and fermented in an immobilized yeast, fluidized bed, fermentation system. The resultant ethanol is recovered in a standard integrated distillation system.

The process design optimizes the energy and water balances by integrating steam producers and users within the plant and recycling process water for washing and dilution. The pentose sugars removed in the water wash are sent to the anaerobic digester to produce a methane-rich gas. The resultant methane-rich gas, recovered lignin, and unconverted cellulose are burned in a fluidized bed boiler to produce high pressure steam for process use.

Those areas of the design in which the technology is not proven are identified as areas requiring additional testing, and research and development. The formulation of a more detailed design for construction and a definitive cost estimate requires specific information for the properties of both the wood and hydrolysate streams through pilot or bench-scale testing. Sterility requirements must be determined and solid separation parameters verified for C_5 recovery, lignin wash, and hydrolysis enzyme recovery units. The effective selection of materials for construction requires more accurate definition of trace components found in the process streams. The effect of process scale-up of the enzyme production and hydrolysis vessels on both air sparging and mixing requirements should be determined.

In addition to the scale-up of bench-scale data, major process uncertainties stem from the effect of inhibitory compounds which may be found in the eucalyptus wood. The process has been designed to reduce the effect of inhibitory compounds by including the water and alkali wash (lignin removal) prior to hydrolysis and enzyme production, and the evaporator system prior to ethanol fermentation.

Wastewater from the plant is recycled as wash water in the first stages of the water/alkali wash and then sent to the anaerobic digester for the production of a methane-rich gas. Only clean condensate from the evaporator is recycled as makeup water to the acid impregnation and hydrolysis sections. Conclusive testing of eucalyptus feedstocks is necessary to determine plant yields and effects of inhibitory compounds, especially if the base case is modified to remove lignin after hydrolysis and/or the evaporator system.

1.2.2 Base-Case Economic Analysis

The capital cost estimate for the enzyme hydrolysis base-case design was generated using budgetary cost estimates from vendors and in-house cost estimates based on equipment specifications. The capital costs of the enzyme-based ethanol plant are shown in Table 1.2-2. The base-case total facility investment is estimated to be $150,624,000, including an allowance for indeterminants. This cost does not include any funds for a process development allowance. Due to the current level of the process design development, a potential increase in the total facilities investment exists and should not be ignored.

The investment analysis of the enzyme-based ethanol plant (EEP) is based on the N^{th} plant in a series of EEPs. The N^{th} plant is assumed to be constructed over a 32-month period from project authorization to commercial operation. The economic analysis is based on a discounted cash flow rate of return (DCFROR) of 15 percent after taxes, a constant dollar basis (1984 dollars) and 100 percent equity financing. The analysis includes the use of the investment tax credit (ITC-10 percent), the energy investment tax credit (EITC-10 percent) and the accelerated cost recovery system (depreciation, ACRS-15, 22, 21, 21, 21 percent) over 5 years. The ACRS is applied to 90 percent of the depreciable plant in accordance with current regulations. The economic basis is shown in Table 1.2-4. No specific alcohol fuel tax credit is taken since it is assumed that this credit is already reflected in the current selling price for fuel ethanol. The base-case design results in a required ethanol selling price of $3.50/gal.

A change in the financial basis to include 25 percent debt at a real interest rate of 8 percent reduces the required ethanol selling price to $3.04/gal. This price is still above the estimated ethanol selling price in Hawaii of $1.80-$2.00/gal and indicates that a better use of the pentose (C_5) and lignin fractions of the wood is required to attract investor financing for this facility.

A comparison of site location for this plant was done to identify the potential changes in the process economics. The comparison site selected was Spokane, Washington. The overall capital costs are essentially unchanged. Savings in construction labor costs and freight charges are offset by the more severe climatic conditions in Spokane. The major operating advantage is in lower wood costs and lower electricity costs. The potential reduction in the base-case ethanol selling price is approximately $0.25/gal.

The summary of operating costs for the base-case ethanol plant and the basis for the investment analysis are shown in Table 1.2-3 and Table 1.2-4, respectively.

1.2.3 Trade-off Studies and Processing Options

The base case design was chosen to utilize current commercial technology wherever applicable. Assumptions for indeterminate by-product markets and technology were not made and, therefore, the design utilized potential by-products internally. The base-case design utilized the pentose fraction of the wood to produce a methane-rich gas in the anaerobic digester. This methane-rich gas and the lignin fraction of the wood are combusted to produce steam for process use. The investment analysis for the base case showed a need to enhance plant revenue. Trade-off studies were performed to evaluate potential economic and technical improvements in process modifications, research and development advances, and by-product sales.

Alternative Uses of By-Products

Trade offs determining the effect of alternative uses of potential by-products on the selling price of ethanol indicate that the sale of the by-products produced from the pentose fractions of the wood and the sale of lignin are the most significant ways to reduce the production cost of ethanol. The most economically attractive uses for the pentose sugars are for the production of ethanol, or the production of furfural. The range of required ethanol prices was determined by analyzing two scenarios. These cases were based on combinations of sensitivity parameters for the additional production of ethanol or furfural from pentose sugars, with and without the sale of lignin. The two scenarios considered were as follows:

- An optimistic scenario which considered a 15-percent decrease in capital costs, the sale of carbon dioxide at $10/ton, a decrease in wood price of $10/dry ton, the assumption of debt, and an increase in the stream factor to 95 percent.

- A pessimistic scenario which considered a 25-percent increase in capital cost, an increase in the wood price of $10/dry ton, the assumption of debt, and a decrease in stream factor to 70 percent.

The ranges of required ethanol selling price between optimistic and pessimistic scenarios for both cases of pentose conversions, with and without lignin sale, are:

Case	Required Ethanol Selling Price	
	W/Lignin Sale(1) ($/gal)	W/O Lignin Sale ($/gal)
Pentose to Ethanol	$1.52-$2.65	$2.08-$3.16
Pentose to Furfural(2)	$1.42-$2.90	$2.14-$3.56

Note: (1) Lignin sold at 15¢/lb, net to plant. (2) Furfural sold at 20¢/lb.

Selling the ethanol or furfural (produced from the C_5 fraction) along with lignin, results in a favorable ethanol selling price under optimistic conditions.

Technical Process Improvements

The sensitivity analyses on the base case are extended by considering potential technical improvements. The areas where viable economic improvement could be achieved through further research and development are increasing enzyme activity and increasing the efficiency of hydrolysis. The recovery of sodium hydroxide has also been identified as an improvement which should be included in any future design. The potential reduction in the required ethanol selling price for these goals are:

Component	Reduction in Ethanol Selling Price (¢/gal)
Enzyme Activity	18
Hydrolysis Efficiency	21
Sodium Hydroxide Recovery	15

These three technical improvements are potentially additive and have a combined effect of potentially lowering the ethanol selling price by approximately $0.50/gal.

The importance of achieving these optimistic research goals is secondary to the sale of by-products, but the combined effect will enhance the attractiveness of the process to private investors.

Some of the less viable technical improvements that may be beneficial are:

- The removal of lignin after hydrolysis
- Elimination of the evaporator
- Elimination of the chip soak pretreatment

These areas of potential economic improvements have certain technical and mechanical uncertainties which require testing before they can be considered commercially viable.

TABLE 1.2-1

ENZYME BASED ETHANOL PLANT - BASIS OF DESIGN

Item	Basis
Plant Capacity	15 MM gal/yr of denatured fuel-grade ethanol
Operating Time	8,000 hr/yr
Location	The Island of Hawaii
Feedstock	Eucalyptus Globulus (Blue Gum): 50 percent Moisture Content: 463,500 ton/yr
Impregnation:	
Sulfuric Acid Conc. in Chip Soak	0.2 wt percent
Retention Time	48 hours
Steam Explosion:	
Pressure	470 psig (saturated)
Retention Time	5 sec at pressure (approximately 40 sec total)
Enzyme Production:	
Enzyme Titre	30 FPU/ml
Residence Time	13 days
Organism	Trichoderma Reesei - Rut - C-30
Hydrolysis:	
Residence Time	48 hours
Glucose Yield	84 wt percent from Steam-Exploded Cellulose
Cellulose Feed Conc	7.0 wt percent
Recovery of Filter Paper Activity	50 percent
Evaporator:	
Type	Five-stage multi-effect evaporator
Concentration of Glucose Achieved	15 wt percent

TABLE 1.2-1 (Continued)

Item	Basis
Fermentation:	
Reactor	Continuous, immobilized bed
Theoretical Yield of Ethanol from Glucose	95 percent

TABLE 1.2-2

ENZYME-BASED ETHANOL PLANT

CAPITAL COST - BASE CASE

Component	Cost (1984 Dollars)
Material and Equipment	54,885,000
Labor	32,915,000
Freight and Tariffs	2,400,000
Total Base Cost	90,200,000
Land Cost	72,000
Engineering and Construction Management	13,312,000
Allowance for Indeterminants	16,876,000
Total Installed Cost	120,460,000
Initial Catalysts and Chemicals	439,000
Startup Expenses (3 months)	12,068,000
Interest During Construction	14,375,000
Working Capital	3,282,000
Total Facilities Investment	150,624,000

TABLE 1.2-3

SUMMARY OF OPERATING COSTS
BASE CASE (1)

Component	Price ($/yr)	Ethanol Cost Contribution (¢/gal)
Raw Materials		
Wood to Process	8,496,072	56.6
Sulfuric Acid	704,900	4.7
Sodium Hydroxide	2,886,000	19.2
Ammonia	940,600	6.3
Wood to Boiler	542,022	3.6
Gasoline	714,304	4.8
Process Chemicals	374,100	2.5
Offsite Chemicals	582,700	3.9
Operating Labor	4,818,000	32.1
Maintenance (Labor & Supplies)	3,660,237	24.4
Electricity	4,644,000	31.0
Insurance and Taxes	1,768,900	11.8
Miscellaneous	849,100	5.7
	30,980,935	206.7

NOTE:

1. 15 x 10^6 gal/yr ethanol production

TABLE 1.2-4

BASIS FOR INVESTMENT ANALYSIS(1)

Component	Amount
Plant Life, years	20
Plant Operation, hr/yr	8,000
Equity, percent	100
Required Return on Investment, percent	15 DCFROR (after tax)
Interest Rate During Construction, percent	8 real
Ethanol Production Rate, Million gal/yr	15
Tax Rate, percent	50
Investment Tax Credit, percent	10
Renewable Energy Tax Credit, percent	10
Depreciation (on 90% of plant)	15, 22, 21, 21, 21
Construction Time, years	2.7
Wood Cost	$39/BDT

Note: (1) The investment analysis is based on a discounted cash flow rate of return (DCFROR) in constant 1984 dollars. The cost of capital (Rate of Return) and interest rates are net of inflation.

1.3 Conclusions

The enzyme hydrolysis process is currently in the bench/pilot scale of development. The design of the base case plant assumes that the laboratory and bench-scale data can be scaled to commercial size. The uncertainties associated with a plant design at this stage of development make definitive economic evaluations questionable.

The investment analysis clearly indicates that by-product credits are necessary to justify continued research and development. Additional economic data for feedstock costs and availability confirmed by-product values and markets are required. The additional market and financial data should then be evaluated against competing fuel grade ethanol feedstocks (i.e., molasses, MSW, bagasse, corn stover, pulp and paper wastes) to determine the relative economic incentive for wood as a feedstock under various economic conditions.

If the market analysis confirms the favorable economics of wood as a feedstock because of the ability to market high value by-products, then continued research and development (R&D) and integrated pilot plant demonstrations would be justified. This feasibility study points out, that, with sufficient by-product credits under optimistic conditions and demonstration of an integrated pilot enzyme hydrolysis plant, competitively priced ethanol can be produced. Substantial R&D is necessary to provide additional economic incentive for the enzyme process. Future R&D goals should be carried out to provide confirmed yield and operating data for commercial plant design.

SECTION 2

INTRODUCTION AND HISTORY OF HYDROLYSIS OF WOOD TO ETHANOL

The conversion of biomass to liquid fuels and chemical feedstocks has been the subject of a great deal of research and development activities in recent years. One of the heavily studied methods for conversion of biomass involves the saccharification of cellulosic materials and fermentation to a fuel-grade ethanol.

Ethanol production by enzymatic fermentation of carbohydrates is a well developed technology. Recently, there has been growing interest in the use of biomass as feedstocks for alcohol production. Cellulose from wood is in abundant supply and is relatively inexpensive, whereas the market for other sources of sugar-containing crops traditionally used in ethanol production may be adversely affected as world population grows and the demand for food supplies increases.

Ethanol production from biomass (renewable cellulosics) by fermentation consists of three distinct stage: hydrolysis, fermentation, and purification. Advances have been made in all three stages, so that today many technically feasible alternatives for alcohol production exist.

Hydrolysis of cellulose consists of breaking down the complex cellulosic polysaccharide to its component sugars, which can be enzymatically fermented to alcohols. Biomass of plant origin contains three major components:

- Cellulose
- Hemicellulose
- Lignin

During hydrolysis, cellulose is converted to glucose (hexoses); and hemicellulose, under milder conditions than that for cellulose, is converted to xylose (pentoses); while lignin remains as a by-product.

The hydrolysis processes that have been, or are being, developed may be classified according to the hydrolyzing agent employed. Acid hydrolysis uses either a dilute mineral acid at high temperatures or a concentrated mineral acid at low temperatures and pressures. Enzymatic hydrolysis uses cellulosic enzymes produced by microorganisms.

Most of the traditional acid hydrolysis processes for production of fermentable wood sugars were developed prior to and during World War II. The better known processes are the Scholler process, which uses dilute sulfuric acid to catalyze the hydrolysis, and the Madison process, which is a continuous percolation acid hydrolysis developed in the United States to improve the economics of the Scholler process. A commercial plant using the Madison process was built in the mid-1940s in

Springfield, Oregon; however, operating difficulties were experienced, and the plant never operated on a commercial basis. Other commercial dilute acid hydrolysis plants were built in Germany and Switzerland in the 1940s, although those plants became uneconomical to operate. During the 1950s, wood hydrolysis plants (some using eucalyptus trees) were built in Russia, where more than 40 plants are still in operation today, and in Brazil, where eucalyptus trees are being used as a feedstock.

Renewed interest in fuel alcohol production as a result of the energy cost escalation of the 1970s has led to several research and development programs in the United States. These programs have concentrated on developing new technologies for making the cellulose-to-ethanol processes economically attractive. Much of the effort has been in the development of enzymatic hydrolysis processes.

Enzymatic hydrolysis of cellulose is a novel development in hydrolysis technology in which high glucose yields of up to 90 percent of theoretical can be obtained. This compares to yields of 50 to 60 percent, for acid hydrolysis. Several processes are currently being developed. No commercial installations have been built to date. The processes being developed can be divided into the following categories:

- Enzyme hydrolysis and subsequent fermentation
- Concurrent enzyme hydrolysis/fermentation, also referred to as simultaneous saccharification and fermentation (SSF)
- Direct microbial processes

An example of each technology is given below:

In the Georgia Tech Process, a portion (10-20 percent) of washed lignocellulose is conveyed to an enzyme fermenter to be used as feedstock for enzyme production. When fermentation is completed, the contents are centrifuged to remove the solids; the enzyme solution is stored in a chilled tank and is ready to be used for enzymatic cellulose hydrolysis. The purified cellulose remaining in this process is hydrolized to fermentable sugars by enzymes. The sugars are subsequently fermented to ethanol.

The Emert Process, also known as the simultaneous saccharification and fermentation (SSF) process, was developed by Gulf Oil and Chemicals at its facility in Pittsburg, Kansas, in the 1970s. In 1979, the technology was donated to the Univeristy of Arkansas by the company. Hence, the process is also known as the Gulf Oil/University of Arkansas Process. The SSF process is continuous in four trains for a total of 12 fermenters. A total residence time of 24 hours is required to complete the reaction. The Georgia Tech and Emert Processes employ the microorganism Trichoderma Reesei for the production of enzymes that can hydrolyze cellulose.

An interesting hybrid process, which combines steam explosion to make the cellulose accessible, enzymatic hydrolysis with T. Reesei, and subsequent fermentation, has been developed by Iotech Corporation of Ottawa, Canada. Pilot plant trials have been conducted. It has been claimed that the process is nearing commercialization.

A thermophilic bacterium that could hydrolyze cellulose rapidly at temperatures up to 60°C has been developed at the University of Pennsylvania. This process is currently under development by Biologies Energy Corporation.

There are research programs aimed at developing the strains of enzyme-producing microorganisms that can efficiently hydrolyze cellulose, reduce hydrolysis times, and produce heat stable enzymes.

of the selection of a specific site, HNEI has provided an estimated cost of $1.00 to $1.50 per 1,000 gallons. Residual fuel oil (No. 6 or Bunker C) is readily available from the local refineries of Chevron and Pacific Resources, Inc. at a 1983 cost of $29 per barrel F.O.B. Hilo Harbor. Propane gas is available via pipeline in the cities of Hilo and Kailua-Kona only. Outside of these areas, facilities may be served by onsite storage tanks. While propane gas cost is dependent upon utilization rate, an average cost for an industrial facility would be approximately $12/MM Btu delivered.

By-product Marketability

In Hilo town, the gas company, GASCO, sells propane to its residential customers. There are no preparation facilities in town. The gas is delivered from Oahu to Hilo for immediate distribution; therefore, direct sale of methane to residential customers would require extensive modification of the existing gas distribution system and would not be feasible. Direct sale of the methane would be possible if an industrial user were available. The economic effect of direct methane is evaluated as a trade-off study.

The pentose sugars which are produced in the enzyme hydrolysis process have a potential value as animal feed. However, the by-product is in direct competition with 311,719 short tons per year of cane molasses, available at $44 per ton. Animal feed by-product sale (see tradeoff) would not be viable in Hawaii.

Furfural and lignin appear to have no local market in the state of Hawaii. Further investigation would be required to determine an export value.

SECTION 4

DESIGN CRITERIA AND BASIS OF DESIGN

A basis of design was determined for the base-case enzyme hydrolysis process. In the selection of the base-case process design, the following general guidelines were used to formulate the design criteria.

1. Maximize production of fuels.

2. Maximize the production of ethanol while minimizing by-product production.

3. Minimize raw material brought to site.

4. Use commercially available equipment, where applicable.

5. Assume technically undeveloped equipment can be scaled to commercial size.

6. Select base-case equipment, based on good engineering practice, which is conservative in actual design but optimistic in expected performance.

7. Maximize process integration.

These criteria were used to select the base case for engineered design and development of a plant cost estimate. Trade-off studies were performed to evaluate the economic and technical effects of process modifications, R&D advances, by-product production, and energy integration on the base case plant costs.

The basis of design for the base case is given in Table 4-1.

TABLE 4-1

BASIS OF DESIGN

General

Plant Capacity: 15 MM gal/yr of denatured fuel-grade ethanol

Operating Time: 8000 hr/yr

Location: The island of Hawaii

Ambient Temperature: 80°F

Ambient Pressure: 14.7 psia

Feedstock: Eucalyptus globulus (blue gum)

Feedstock Composition:

	Wt %
Extractives	1.3
Lignin	21.9
Celluose	51.3
Glucomannan	1.4
Glucuronoxylan	19.9
Other Polysaccharides	3.9
Residual Constituents	0.3
	100.0

Moisture Content:	50%
Chip Size:	Nominal 3/4-in. size from chipper
Feedstock and product storage:	Wood: 14 days Chemicals: 14 days Ethanol Product: 14 days

Impregnation (100)	
H_2SO_4 concentration in chip soak:	0.2%
Liquid to dry wood in soak tank:	4:1
Mositure content retained in chips:	60%
Temperature in soak tank:	109°F
Retention time:	48 hr
Steam Explosion and Washing (200)	
Pressure:	470 psig (saturated)
Retention time:	5 sec at pressure (approximately 40 sec total)
Steam required:	680 lb/ton wet wood (steam recovery not included)
Furfural yield:	0.005 lb furfural/lb dry wood
Pseudo-lignin yield:	0.023 lb pseudo-lignin/lb dry wood (generated from pentose sugars)
Flash I pressure:	60 psig into pressure cyclone for heat recovery
Flash II pressure:	5 psia
Countercurrent Water/Alkali Wash:	
Percent hemicellulose sugars solubilized:	80%
Percent recovery of solubilized hemicellulose sugars:	85%
NaOH Requirement:	0.17 lb NaOH/lb lignin dissolved
Lignin solubilized:	75%
Percent recovery of solubilized lignin:	85%

Enzyme Production (300)	
Enzyme productivity:	96 FPU/1-hr
Enzyme titre:	30 FPU/ml
Residence time:	13 days
Cellulose feed concentration:	150 g/1
Organism:	Trichoderma - Rut - C-30
Enzyme specific activity:	0.6 FPU/mg protein
Supplemental media:	15 g/1 corn steep liquor (dry basis) 23 g/1 NH_4OH (as nutrient and pH reguation)
Temperature:	90°F
pH:	4.8
Hydrolysis (500)	
Residence time:	48 hr
Glucose yield:	84 wt % from steam exploded cellulose
Enzyme dosage:	25 FPU/gm solids
Cellulose feed concentration:	7.0 %
Cellulose wt conversion to glucose:	1.1 lb glucose/lb celluose hydrolyzed
Temperature:	122°F
pH	4.8
Recovery of filter paper activity:	50%
Recovery of C_1 C_X enzyme:	50% (for material balance)
Recovery of β-glucosidase enzyme:	20%(for material balance)

TABLE 4-1 (Cont)

Evaporation Prior to Fermentation (500)	
Type of evaporator:	5-stage multi-effect evaporator
Concentration of glucose achieved:	15 wt %
Feed temperature:	180°F
Fermentation (600)	
Reactor configuration:	Continuous, immobilized bed
Theoretical yield of ethanol from glucose:	95%
Temperature:	85°F
Initial pH:	4.0
Ethanol Recovery (700)	
A. Beer Still	
Temperature Feed	150°F
Temperature Bottom	238°F
Temperature Overhead	182°F
Pressure Bottom	24 psia
Pressure Top	17 psia
Overhead Composition	Azeotrope
Bottom Composition	Ethanol 0.0001 wt %
B. Anhydrous Distillation	
Entrainer	Cyclohexane
Ethanol Product Purity	99.5 vol % ethanol
Anaerobic Digestion (800)	
Product	Methane
Yield	5.6 scf CH_4/lb degradable organics; 80% reduction of COD
Composition (by volume)	60% Methane, 40% Carbon Dioxide

<u>Lignin Boiler</u> (900)

pH at which all lignin is precipitated:	9.0
Percent solids leaving lignin centrifuge:	38%
Type of boiler:	Fluidized bed
Supplemental Fuel:	1) Methane-rich gas 2) Wood chips

SECTION 5

PROCESS DESCRIPTION

5.1 PROCESS PLANT

The process plant descriptions are given below by plant section number. Process flow diagrams for each section are shown in Figures 5.1-1 to 5.1-17.

Section 1000 - Feedstock Handling

Chipped wood is received from the eucalyptus tree farm in 50-ft chip vans. These vans are dumped into the wood chip unloading pit (G-1004). The nominal 3/4 in. chips are then removed from the pit via drag chains (W-1001, W-1002), cleaned of metallics (G-1006), and oversize material is removed in a feed scalper (G-1005). Fourteen days storage of chipped wood is provided in two stacked piles. The piles are stacked by a double wing belt stacker conveyor (W-1003). The chips are recovered by two traveling cantilever scraper reclaimers (W-1005A,B) and transferred to surge bin (Q-1002) via belt conveyors (W-1006, W-1008). Tramp metal is removed from the chips in a magnetic separator (G-1007).

The wood is then graded to remove rocks and sand. Triple deck chip screens (G-1010A,B) are provided to separate the chips into three fractions - undersize, nominal size, and oversize. The nominal-sized chips are conveyed to the main feed conveyor (W-1012) and the boiler fuel transfer conveyor (W-1013) by a belt conveyor (W-1011). The oversized chips are dropped into a flotation stone trap (G-1008), hogged in a grinder (G-1009), and mixed with the nominal chip fraction. The undersize chips are passed over a fines recovery screen (G-1013) to remove sand and grit. The clean fines are transferred to the boiler fuel transfer conveyor (W-1013).

The main feed conveyor (W-1012) transports the processed chips to the pretreatment inclined conveyor feeder (W-101), Section 100. Chip weight is totalized on the main feed conveyor (W-1012) by the belt conveyor scale (G-1011). Samples are taken for laboratory analysis to determine the feed chip chemical breakdown.

Section 100 - Pretreatment

In the pretreatment section, the wood chips are soaked with 0.2 wt percent sulfuric acid for 48 hours. This pretreatment step is included to enhance the desireable effects of the steam explosion treatment at lower steam explosion severities. The wood chips are fed via the main feed belts (W-101, W103A,B) to six impregnation vessels (M-101A-F). Each vessel has capacity to soak 8 hours of wood chip feed for approximately 48 hours. A sulfuric acid solution is continually circulated through the pretreatment vessels using the acid recycle pumps (P-101A-F). At the end of the soak cycle, the sulfuric acid solution is drained to the next impregnation vessel for wood chip filling. The chips are then surface-washed with recycle waters and screwed from the impregnation vessels using standard screw bottom assemblies (W-104A-F) which are fitted to the bottom of the vessels. Air cannons (V-105A-X) are supplied to ensure chip movement from the vessel. The impregnated chips are then transported to the steam explosion, Section 200, by conveyor screws (W-105A-F) and a central belt conveyor (W-106).

The acid pretreatment section is an area where capital and operating cost could be reduced either by alternate methods of acid soaking (i.e., open pits, presized soaking, acid washing, pressure soak, etc) or by the total elimination of the soak system. The reasons for its inclusion in the base case are discussed in the trade-off study (Section 8.8).

Section 200 - Steam Explosion/Wash

The acid-soaked wood chips are loaded into the steam explosion feed bins (M-203A-D) by a vibrating rotary feeder (W-201). The feed bins provide surge capacity for the intermittent loading of the steam explosion guns (V-203A-D). The mechanism of steam explosion is to disrupt the cellulose microfibrils, detach the lignin, increase the surface area of cellulose available for enzymatic attack, and increase the susceptibility of the cellulose to hydrolysis by altering the cellulose crystalline structure. Four explosion guns, each containing 30 cubic feet, are required to process the wood feed. Each steam explosion gun is fired once every 22.5 seconds. A single steam explosion gun cycle consists of loading the gun with wood chips; raising the gun internal temperature and pressure to 464°F, 470 psig by high pressure (HP) steam injection; cooking the chips for 5 seconds at temperature; and explosion of the chips into the MP flash vessel (M-201). The explosion consists of forcing the softened chips out of the steam explosion vessels by rapidly reducing the systems pressure. The energy of steam explosion is released as 60 psig steam in the MP flash vessel where the flash steam is recovered for process use. The flash vessel is of sufficient size to dampen the intermittent steam surges from all four guns. The defibrated exploded pulp is then cooled by vacuum flash in the vacuum flash vessel (M-202). A portion of the beer still (A-701), Section 700, bottoms is recycled to the vacuum flash to wash the solids fraction (pulp) from the vessel. The flashed vapors are condensed in a knockback condenser (T-201). A vacuum pump (R-201) is installed to remove noncondensibles from the flash vessel to maintain vacuum conditions.

The cooled pulp is now processed in the counter-current water/alkali wash (V-201). The counter-current wash system consists of a traveling belt filter unit. This washer is limited to a 10 wt percent solid feed so that liquid may be properly distributed along the length of the wash unit. To obtain this consistency, the pulp in the vacuum flash vessel (M-202) is diluted with recycle wash waters from the counter-current wash (V-201). The purpose of the counter-current wash unit is to remove the soluble hemicellulose sugars and extract the lignins with sodium hydroxide. In the first wash stages, the pulp is washed with beer still (A-701), Section 700, bottoms product. This is followed by a clean wash with process makeup water. The water wash is essential to remove the water soluble degradation products (phenolics, wood derivatives, etc) that are inhibitors of yeast fermentation and the production and activity of the enzyme complex.

The pulp then continues on the wash (V-201) belt where a caustic solution is used to solubilize and extract the lignin from the cellulose/lignin complex. It is also possible to extract and recover lignin after hydrolysis. This option was not used in the base case for the following reasons:

1. Higher solids flow on conveying equipment would increase handling costs

2. Increased solids in the enyzme fermenters and hydrolysis fermenters could create mixing problems

3. Enzyme recovery yield could be lowered because of enyzme adsorption on lignin.

4. The hydrolysis rate and overall glucose yield is decreased because the lignin associated with the cellulose is still expected to shield or block a small portion of the cellulose from enzyme attack.

5. Lignin recovered by extraction is a more desirable reactive lignin than lignin recovered after hydrolysis.

The amount of caustic required to solubilize the lignin is 0.17 lb sodium hydroxide (NaOH) per lb lignin dissolved. It was assumed that 75 percent of the lignin would be solubilized at this caustic rate. Confirmed data is required to verify these assumptions.

The solubilized lignin stream is pumped to the boiler island where it is treated with sulfuric acid to precipitate the lignin. The lignin is then concentrated as boiler feed in the lignin centrifuge (G-904). The unsolubilized lignin which leaves the caustic wash (V-201) passes through the enzyme fermentation and hydrolysis sections and is ultimately recovered in the hydrolysis centrifuge (G-404), Section 400. The recovered lignin is then added to the lignin precipitate to be burned as boiler fuel.

Section 300 - Enzyme Production

The total hydrolysis makeup enzyme requirement is provided in the enzyme production section. The system design employs a fed batch enzyme fermentation, modeled after experimental data developed at the University of California, Berkeley. The organism used is Trichoderma Rut-C-30. A 13-day residence time is required to achieve a final titre of 30 Filter Paper Units/ml (FPU/ml). Fermenters are batch-fed to achieve a total cellulose concentration equivalent to a 15 wt % (150 g/l) cellulose batch operation.

The enzyme production section consists of two trains of three batch-fed fermenters. Each fermenter train has a total of 13 days residence time, with each individual fermenter maintaining 104 hours of residence. The initial charge is conveyed (W-401 Section 400, W-302) into the first fermenter (M-301A,B), where it is diluted with sterile water, adjusted for pH, fed nutrient, and inoculated with organism. Inoculation rate is assumed at 1 percent of the initial charge. Reduced fermentation times have been reported at higher inoculation rates. Additional research, including recycling of organism, could reduce cost in this area. This is presented in trade-off study 8.5. The fermenter vessels are 18-ft-dia by 60-ft-high, cone bottom, vertically stirred vessels with supplemental air sparging. The sterile air is supplied by air compressors (R-301A,B), which are driven by backpressure steam turbines. The mixing and air sparge rates were optimized using standard general mass transfer techniques. For economic reasons the rates calculated for commercial-type operation are not as high as rates used in bench-scale testing. Yield data and mass transfer behavior must be demonstrated on a pilot-size fermenter vessel so that fermenter vessel and mixing requirements can be adequately determined for a commercial-size facility. The reaction heat is removed from the fermenter vessels by external circulation heat exchangers (T-301A,B) cooled with chilled water. After 104 hours of residence time in the first fermenter vessel (M-301A,B), the partially fermented broth is pumped to the second fermenter (M-302A,B). Here the fermentation is continued for an additional 104 hours in a batch-fed fermentation vessel. The remaining cellulose requirement is added during the second 104 hours of fermentation. The broth is then pumped to the final enzyme fermenters (M-303A,B), where the remaining 10 percent of enzyme production is performed. The final fermentation continues in a batch mode; here no provision is made for intermittent feeding of cellulose or cooling of the fermentation broth. The third fermenter contains sufficient enzyme to supply 6.5 batch additions to the hydrolysis reactors (L-401A,B to L-406A,B) Section 400. The first batch addition is drawn from the fermenter (M-304) after 52 hours of fermentation with subsequent inoculations drawn every 8 hours thereafter. This method of inoculation eliminates the need for an intermediate enzyme storage tank and the reduction of enzyme activity which would occur during storage. It can readily be seen that the total fermentation broth does not attain the full 312 hours of fermentation; however, the small change in enzyme titre over the last 52 hours of fermentation can justify this inoculation method.

Corn steep liquor (30 g/l at 54-percent solids) is added to supplement the media nutrients. In addition to the corn steep liquor, a supplemental nitrogen source and pH regulator, ammonia, is added in the amount of 23 g/l. This amount was based on the quantity required to maintain nitrogen balance for enzyme production and cell synthesis.

Section 400 - Hydrolysis

Fermentable sugars are formed by enzymatic attack of the cellulose complex in the hydrolysis section. As with any biological process, the sugar yield is dependent on substrate concentration, reaction time, and also any inhibitory component present. Data were obtained from various sources which indicated that glucose released during hydrolysis inhibits the enyzme complex. A buildup of the intermediate product, cellobiose and other reducing sugars, occurs without a complete conversion to glucose. To obtain high yields (90 percent or higher), residence time of about 48 hours and final glucose concentrations of 5 percent or less are required. These yields can be obtained in laboratory-scale units; however, in commercial scale plants with enzyme recovery, these yields may not be possible.

The primary data were extracted from a Master's thesis by Stephen T. Orichowskyj (University of California, Berkeley). These data were generated from steam exploded corn stover. This information was used in the base case since information on steam-exploded eucalyptus is not readily available. The glucose concentration in hydrolysis was chosen as 7 wt percent. At this concentration, yields of 84 percent of cellulose conversion in 48 hours are anticipated. The temperature and pH of hydrolysis are specified as 122°F and 4.8, respectively.

Enzyme recovery is included as part of the base-case design. The first two enzymes, cellobiohydrolase and endoglucanase (C_1C_X), in the overall enzyme complex have a high affinity for the insoluble cellulose substrate and are easily recovered by adsorption onto fresh substrate. The third major enzyme, cellobiose (beta-glucosidase) acts upon the soluble cellobiose dimer and is, therefore, not readily adsorbed onto fresh cellulose substrate. It is reported that beta-glucosidase enzyme is associated with the cell wall of the mycelia; therefore, if the cells are added with the enzyme broth to hydrolysis, the beta-glucosidase is in sufficient supply. The base case assumes a 50 percent recovery of enzyme filter paper activity.

The hydrolysis section was designed to reduce the possibility of contamination when recovering enzyme. This was accomplished by restricting enzyme recycle to the hydrolysis reactor from which it was recovered. The hydrolysis section consists of 12 hydrolysis reactors (L-401A,B to L-406A,B). In addition to the hydrolysis reactors, there are two enzyme recovery vessels (M-401A,B). The hydrolysis reactors and enzyme recovery tanks are standard pulp and paper high density stock tanks with bottom mixing. Agitation requirements are based on requirements for moving a thick pulp and paper stock. These reactors are cycled in pairs on a 48-hour turnaround. A complete hydrolysis cycle can be described by considering one pair of hydrolysis reactors (L-401A,B) in sequence with the enzyme recovery tanks (M-401A,B). The washed cellulose feed is conveyed to the enzyme recovery tanks (M-401A,B) via the hydrolysis feed conveyor (W-402A). During this time period, the hydrolysis reactors (L-401A,B) are drained. About 30 percent of the hydrolysate is sent to the hydrolysis centrifuge (G-404) where unconverted cellulose and residual solids are removed. The solids are conveyed via water wash feed conveyor (W-403) to the counter-current water wash (V-401). The washed solids are sent to the boiler while the wash liquids are forwarded to the evaporator. The filtrate from the hydrolysis centrifuge (G-404) is recycled to the enzyme recovery tanks (M-401A,B). The remaining 70 percent of hydrolysate from the hydrolysis reactors (L-401A,B) is sent directly to the enzyme recovery tanks (M-401A,B) for enzyme adsorption. The empty hydrolysis reactors are cleaned and then fed with fresh feed cellulose. About 50 percent of the feed cellulose is sent to the enzyme recovery tanks (M-401A,B) and 50 percent to the hydrolysis reactors (L-401A,B). While the hydrolysis reactors are filling, makeup evaporator water and fresh enzymes are added. The enzyme recovery tanks (M-401A,B) are then drained for an 8-hour cycle to provide constant feed to the evaporators (500 section). The cellulose feed which was conveyed to the enzyme recovery tanks (M-401A,B) is separated from the hydrolysate in the hydrolysis recycle centrifuges (G-403A,B) and returned to the hydrolysis reactors (L-401A,B). The enzyme recovery tanks (M-401A,B) are then prepared for the next cycle.

The enzyme hydrolysis is based entirely on laboratory-scale data. Testing and demonstration is required to design a commercial-size facility. Specific areas where additional research is necessary are enzyme recovery data, hydrolysis yield and rate data, identification of inhibitors, and sterility requirements of hydrolysis. This research data should be obtained under experimental conditions which simulate commercial scale operating conditions so that scale-up effects can be minimized.

Section 500 - Evaporation

The hydrolysate stream leaving the 400 section contains 5.7 wt percent fermentable sugars. An evaporation step is included to increase the glucose concentration to 14.7 wt percent. The primary purpose of the evaporation step is to provide a fermentation feedstock which is readily adaptable to current commercial fermentation and alcohol recovery technologies. The evaporation system also provides a clean condensate for inplant water recycle. Elimination of the evaporation system could reduce capital costs and enhance the plant steam balance. Pilot plant testing of low concentration sugar fermentations, low alcohol concentration distillation methods, and the effects of "dirty" water recycle on plant operability, are required before the evaporation step can be removed.

The evaporation system consists of a standard five-stage, multi-effect evaporator. The glucose stream enters the first evaporation stage where it is vaporized against boiler low pressure (LP) steam. The generated vapor in each successively lower pressure evaporation stage is used to vaporize additional water from the glucose stream. The vapors from the final stage of evaporation are condensed in the evaporator surface condenser (T-501A,B). The concentrated glucose leaves the lowest pressure stage at 14.7 wt percent glucose concentration.

Section 600 - Fermentation

The fermentable sugars are converted to ethanol in a continuous fermentation system. The base-case design uses a process which is similar to a process promoted by Kyowa Hakko Kogyo Co., Ltd. This process is based on yeast immobilized in resin beads which are packed in large stainless steel fermenters (L-601A-C, L-602A-C). The process flow diagram (Figure 5.1-6) is a representation of this fermentation process.

Typical conversions obtained in this type of system approximate the theoretical maximum. Test results indicate that about 95 percent of theoretical conversion is possible when operating at 86°F to 90°F. For mass balance purposes, all of the sugars are assumed to ferment at the same rate, with 2 percent of the sugar used for yeast production and 3 percent for by-product formation. It is anticipated that 4-month operating runs are obtainable before system sterilization is required. A clean-in-place system is incorporated in the plant design to provide the intermittent sterilization requirements.

The clean, concentrated hydrolysate is fed to the fermenters through the fermenter feed cooler and chiller (T-601, T-604), where the hydrolysate is cooled to 85°F. Three trains of two-stage fermenters are incorporated. A small quantity of oxygen is added by an air sparge compressor (R-601) to maintain yeast productivity. The fermenters are maintained below 90°F by internal fermenter coolers. The plant cooling water supply temperature is too high to provide adequate cooling of the fermenters: therefore, a package chilled water system is provided. Equipment for replacement and recovery of immobilized yeast beads is included as part of the fermentation system. Approximately 80 percent of the carbon dioxide can be recovered as saleable by-product. Recovery equipment is not included in the base case design.

Section 700 - Distillation

The beer still system is designed to produce 190-proof ethanol product. The column feed (clarified fermenter effluent) is pumped from the fermenters to the beer still feed tank (M-705). Recycles from the fusel oil decanter (M-703), vent system and CO_2 scrubbers (A-1202, A-1201) Section 1200, and degasser drum vent condenser (T-727) are also collected in this tank. The tank is designed for a 15-minute holdup. Its effluent is pumped by the beer still feed pump (P-716) to the beer still feed preheater (T-701), where the beer still feed is heated to 150°F. The heat source for this exchanger is a portion of the beer still overhead vapor. After preheating, the feed enters the degasser drum (M-706) where dissolved carbon dioxide is released. This vapor is cooled in the degasser drum vent condenser and sent to the vent recovery system. The degasser drum is elevated above the beer still feed location. The liquid feed passes from the drum to the beer still by gravity flow.

The beer still is an atmospheric column in which conventional distillation is used to produce an azeotropic ethanol/water product. Two types of trays are employed in this column. The soluble solids present in the feed tend to form deposits on the underside of the trays below the feed point. Ripple trays have been specified because their self-cleaning features virtually eliminate these deposits. Since fouling is not anticipated above the feed location, valve trays have been specified for this section.

Steam from the dirty LP steam header is used to reboil the column by direct steam injection. Direct injection of steam will reduce the plating and corrosion problems associated with a beer still reboiler.

Additional reboiler steam requirements are provided by boiler LP steam in the beer still reboiler (T-708).

Above the main feed tray there are four fusel oil sidedraw nozzles. Fusel oils accumulate in the column and are removed from the column side draws.

There is a pasteurizing section of five trays at the top of the beer still. A 95 volume percent ethanol product is removed as a sidedraw from the tray below this section and sent to the anhydrous system via anhydrous column feed pump (P-707A&B).

The beer still overhead vapor is used to preheat the feed in the beer still preheater. The condensed vapors are sent to the beer still reflux drum (M-701). Noncondensables are vented to the beer still vent condenser (T-726).

The beer still trim condenser (T-702) is mounted directly above the beer still reflux drum. Thus, the vapors rising from the drum are condensed and returned to the drum. The small condensate stream produced has a high ethanol content and is sent to fusel oil storage tank (M-713).

The fusel oil sidedraw streams are combined with evaporator condensate before entering the fusel oil cooler (T-712). Here, the stream is cooled to 100°F. This sequence allows the ethanol in the sidedraw to be absorbed by the water. On cooling, two immiscible liquid phases are formed. These are introduced into the fusel oil decanter (M-703) where the water/ethanol mixture forms a lower layer and the fusel oil product is the upper layer. The fusel oil product flows into the fusel oil storage tank (M-713). The water/ethanol product is returned to the beer still feed tank. The fusel oil decanter is elevated to allow for gravity flow of both product streams.

The anhydrous system consists of two distillation columns. The feed enters the anhydrous column (A-702). In this column, cyclohexane is used as an entrainer to separate the near azeotropic ethanol (95 percent, 5 percent water), produced in the beer still (A-701). Entrainer in a mixture with ethanol and water is constantly returned to the column via the anhydrous column recycle pump (P-715) from the anhydrous column recycle drum (M-707). The bottom product is 99.98 percent ethanol. This product is cooled to 100°F against cooling water in the product cooler (T-704) before being sent to storage. The column vapor is generated in a thermosyphon reboiler (T-719) using LP process steam as the heating medium.

The overhead from the anhydrous column is condensed by cooling water in the anhydrous column overhead condenser (T-715), then returned to the anhydrous column reflux drum (M-708) for reflux. A small liquid purge stream is withdrawn from this drum. This purge is cooled in the anhydrous purge cooler (T-713) and combined with the purge from the recovery column (A-703) before entering the wash solvent purification system.

The overhead product from the anhydrous column is a three-component azeotrope. It is withdrawn from a liquid drawoff tray, just below the pasteurization section of the column. This bubble-point liquid is cooled to 100°F in the decanter feed cooler (T-711) before entering the anhydrous system decanter (M-702). The upper, ethanol/cyclohexane-rich phase, is combined with the cyclohexane-rich product drawoff from the recovery column (A-703). The combined stream flows to anhydrous column recycle drum (M-707). The water-rich lower phase serves as feed to recovery column (A-703).

The recovery column (A-703) separates the remaining ethanol and cyclohexane from the water. Column vapors are generated by LP steam in the recovery column reboiler (T-709).

The overhead from this column is condensed in the recovery column condenser (T-703) and flows to recovery column reflux drum (M-709). Reflux pump (P-703A&B) withdraws the condensed overhead from the reflux drum and returns it to the recovery column. A small purge stream is taken from the reflux and combined with the anhydrous column reflux purge which is cooled in the anhydrous column purge cooler (T-713). The combined streams flow into the solvent purification system.

The recovery column overhead product is withdrawn from a plate below the pasteurization section. It is then combined with the ethanol/cyclohexane-rich phase from the anhydrous system decanter (M-702) and stored in the anhydrous recycle drum (M-707). The total flow is eventually returned to the anhydrous column (A-702).

The anhydrous column recycle drum (M-707) serves as a collecting tank for process streams rich in cyclohexane. These streams include the decanter top phase, recovered solvent, the liquid sidedraw from the recovery column (A-703), and control streams from the reflux of each column.

Section 800 - Anaerobic Digestion

The anaerobic digestion system consists of a Barcardi Corporation anaerobic filter (L-801) and its associated equipment. The digester is a 3.5-million-gallon anaerobic filter packed with corrugated plastic media which supports a film of active microorganisms. The organic components from the feed stream are biologically converted to a 50 to 60 percent methane gas stream which is burned as boiler fuel. The COD (Chemical Oxygen Demand) reduction is about 80 percent. The wastewater from the plant first enters the digester hold tank (M-801). The hold tank is included for feed surge control. The wastewater is then cooled in the digester feed cooler (T-801) prior to entry into the anaerobic digester (L-801). The liquid overflow from the digester is sent to the wastewater treating system for final treating. The methane rich gas is compressed in the methane gas compressor (R-801) for use as boiler fuel. A gas storage sphere (M-803) is included for inplant storage capacity.

TABLE 5.1-1

MATERIAL BALANCE COMPONENTS

(Section 100 - Pretreatment)

Stream Number	102	103	104
Stream Name	Makeup Sulfuric Acid	Feed to Steam Explosion Bins	Recycle H_2SO_4/H_2O
	lb/hr	lb/hr	lb/hr
Component			
Water	-	81,926	144,680
Cellulose	-	27,939	-
Glucose	-	-	-
Other Hexose	-	763	-
Pentose	-	10,838	-
Lignin	-	11,927	-
Ethanol	-	-	-
Degraded Pentose	-	-	-
Furfural	-	-	-
Solubles	-	871	-
Insolubles	-	2,124	-
Sulfuric Acid	156	156	275
Sodium Hydroxide	-	-	-
Sodium Sulfate	-	-	-
Enzyme	-	-	-
Mycelia	-	-	-
Nutrients	-	-	-
Ammonium Hydroxide	-	-	-
Carbon Dioxide	-	-	-
Air	-	-	-
Total	156	136,544	144,955
Temperature, °F	85	117	117

TABLE 5.1-1 (Cont)

(Section 200 - Steam Explosion/Wash)

Stream Number	201	202	203	204	205	207
Stream Name	MP Flash Vapor to Beer Still	Vacuum Flash Feed	Vacuum Flash Condensed Vapor	10% Solids Waterwash Feed	Process Water to Water/NaOH Wash	Feed to Anaerobic Digestion
	lb/hr	lb/hr	lb/hr	lb/hr	lb/hr	lb/hr
Component						
Water	25,678	102,801	18,449	554,728	7,501	161,860
Cellulose	-	27,939	-	30,778	-	207
Glucose	-	-	-	104	-	486
Other Hexose	-	763	-	995	-	1,248
Pentose	-	9,309	-	10,785	-	8,869
Lignin	-	11,927	-	13,162	-	1,212
Ethanol	-	-	-	-	-	-
Degraded Pentose	-	1,256	-	1,395	-	65
Furfural	-	273	-	302	-	238
Solubles	-	871	-	1,004	-	938
Insolubles	-	2,124	-	2,359	-	107
Sulfuric Acid	-	156	-	148	-	-
Sodium Hydroxide	-	-	-	-	-	-
Sodium Sulfate	-	-	-	82	-	385
Enzyme	-	-	-	158	-	737
Mycelia	-	-	-	-	-	-
Nutrients	-	-	-	16	-	74
Ammonium Hydroxide	-	-	-	-	-	8
Carbon Dioxide	-	-	-	-	-	-
Air	-	-	-	-	-	-
Total	25,678	157,419	18,449	616,016	7,501	176,434
Temperature, °F	308	308	162	162	80	129

TABLE 5.1-1 (Cont)

(Section 200 - Steam Explosion/Wash)

Stream Number	209	210	211	212	213	214	215
Stream Name	Alkali/H_2O Feed to Wash	Lignin Feed to Boiler	Cellulose to Hydrolysis and Enzyme Production	Steam Exploded Wood to Flash	Alkali/Water Wash Recycle	Vacuum Flash Vessel Vent	Stm to Stm Explosion
	lb/hr	lb/hr	lb/hr	lb/hr	lb/hr	lb/hr	lb/hr
Component							
Water	120,941	120,949	116,063	128,479	419,858	-	46,553
Cellulose	-	-	27,939	27,939	2,799	-	-
Glucose	-	-	-	-	11	-	-
Other Hexose	-	35	209	763	93	-	-
Pentose	-	430	2,549	9,309	990	-	-
Lignin	-	6,957	3,956	11,927	1,197	-	-
Ethanol	-	-	-	-	-	-	-
Degraded Pentose	-	-	1,256	1,256	127	-	-
Furfural	-	35	6	273	28	-	-
Solubles	-	111	20	871	92	Inerts	-
Insolubles	-	-	2,123	2,124	215	-	-
Sulfuric Acid	-	-	-	156	19	-	-
Sodium Hydroxide	1,443	1,210	213	-	-	-	-
Sodium Sulfate	-	28	6	-	9	-	-
Enzyme	-	-	-	-	17	-	-
Mycelia	-	-	-	-	-	-	-
Nutrients	-	-	-	-	-	-	-
Ammonium Hydroxide	-	-	-	-	-	-	-
Carbon Dioxide	-	-	-	-	-	-	-
Air	-	-	-	-	-	-	-
Total	122,384	129,755	154,340	183,097	425,455		46,553
Temperature, °F	80	101	94	308	129	162	464

TABLE 5.1-1 (Cont)

(Section 200 - Steam Explosion/Wash)

Stream Number	216
Stream Name	Caustic Dilution Water
	lb/hr
Component	
Water	119,498
Cellulose	-
Glucose	-
Other Hexose	-
Pentose	-
Lignin	-
Ethanol	-
Degraded Pentose	-
Furfural	-
Solubles	-
Insolubles	-
Sulfuric Acid	-
Sodium Hydroxide	-
Sodium Sulfate	-
Enzyme	-
Mycelia	-
Nutrients	-
Ammonium Hydroxide	-
Carbon Dioxide	-
Air	-
Total	119,498
Temperature, °F	80

TABLE 5.1-1 (Cont)

(Section 300 - Enzyme Production)

Stream Number	304	305	306	307	308	309	310
Stream Name	Sterile Water	Total Nutrients Added	Vent	Air for Sparging	Enzyme Production Product Stream	HP Steam (615 psia)	LP Steam (65 psia)
	lb/hr	lb/hr	lb/hr	lb/hr	lb/hr	lb/hr	lb/hr
Component							
Water	3,947	1,067	-	-	14,269	91,909	91,909
Cellulose	-	-	-	-	208	-	-
Glucose	-	-	-	-	-	-	-
Other Hexose	-	-	-	-	17	-	-
Pentose	-	-	-	-	203	-	-
Lignin	-	-	-	-	315	-	-
Ethanol	-	-	-	-	-	-	-
Degraded Pentose	-	-	-	-	100	-	-
Furfural	-	-	-	-	-	-	-
Solubles	-	-	Inerts	-	2	-	-
Insolubles	-	-	-	-	169	-	-
Sulfuric Acid	-	-	-	-	-	-	-
Sodium Hydroxide	-	-	-	-	-	-	-
Sodium Sulfate	-	-	-	-	31	-	-
Enzyme	-	-	-	-	742	-	-
Mycelia	-	-	-	-	223	-	-
Nutrients	-	223	-	-	74	-	-
Ammonium Hydroxide	-	341	-	-	104	-	-
Carbon Dioxide	-	-	1,440	-	-	-	-
Air	-	-	-	111,708	-	-	-
Fusel Oils	-	-	-	-	-	-	-
Other Volatiles	-	-	-	-	-	-	-
Gasoline	-	-	-	-	-	-	-
Ash	-	-	-	-	-	-	-
Total	3,947	1,631	1,440	111,708	16,457	91,909	91,909
Temperature, °F	60	82	82	120	82	750	-

TABLE 5.1-1 (Cont)

(Section 400 - Hydrolysis)

Stream Number Stream Name Component	401 Pretreated Cellulose Feed to Hydrolysis Reactor lb/hr	402 Pretreated Cellulose Feed to Enzyme Recovery Tanks lb/hr	403 Reactor Product to Enzyme Recovery Tanks lb/hr	404 Recyle from Hydrolysis Centrifuge lb/hr
Water	53,408	53,408	311,736	138,166
Cellulose	12,856	12,856	7,948	-
Glucose	-	-	21,840	9,793
Other Hexose	96	96	111	49
Pentose	1,173	1,173	1,348	597
Lignin	1,820	1,820	7,581	-
Ethanol	-	-	-	-
Degraded Pentose	578	578	2,407	-
Furfural	3	3	6	3
Solubles	9	9	10	5
Insolubles	977	977	4,071	-
Sulfuric Acid	-	-	-	-
Sodium Hydroxide	98	98	-	-
Sodium Sulfate	3	3	204	90
Enzyme	-	-	746	627
Mycelia	-	-	405	-
Nutrients	-	-	71	32
Ammonium Hydroxide	-	-	99	44
Carbon Dioxide	-	-	-	-
Air	-	-	-	-
Total	71,021	71,021	358,583	149,406
Temperature, °F	94	94	122	122

TABLE 5.1-1 (Cont)

(Section 400 - Hydrolysis)

Stream Number	405	406	407	408
Stream Name	Feed to Hydrolysis Recycle Centrifuge	Solids Recyle to Hydrolysis	Hydrolysis Product to Evaporator	Reactor Product to Enzyme Recovery Tank and Centrifuge
Component	lb/hr	lb/hr	lb/hr	lb/hr
Water	503,407	144,040	395,773	474,398
Cellulose	20,804	20,804	207	12,095
Glucose	31,633	9,054	24,101	33,246
Other Hexose	256	55	209	168
Pentose	3,118	675	2,539	2,051
Lignin	9,401	9,401	198	11,536
Ethanol	-	-	-	-
Degraded Pentose	2,985	2,985	63	3,663
Furfural	12	6	6	9
Solubles	24	5	20	16
Insolubles	5,048	5,048	106	6,195
Sulfuric Acid	-	-	-	-
Sodium Hydroxide	-	-	-	-
Sodium Sulfate	472	102	383	310
Enzyme	1,373	742	737	1,484
Mycelia	405	405	12	628
Nutrients	103	34	74	108
Ammonium Hydroxide	143	48	103	151
Carbon Dioxide	-	-	-	-
Air	-	-	-	-
Total	579,184	193,404	424,531	546,058
Temperature, °F	120	120	115	122

TABLE 5.1-1 (Cont)

(Section 400 - Hydrolysis)

Stream Number Stream Name Component	409 Hydrolysis Centrifuge Feed lb/hr	410 Hot Evaporator Feed From Hydrolysis lb/hr	411 Hydrolysis Centrifuge Solid to Water Wash lb/hr	412 Process Water to Wash lb/hr
Water	162,662	395,773	24,495	36,714
Cellulose	4,147	207	4,147	-
Glucose	11,406	24,101	1,613	-
Other Hexose	58	209	9	-
Pentose	703	2,539	106	-
Lignin	3,956	198	3,956	-
Ethanol	-	-	-	-
Degraded Pentose	1,256	63	1,256	-
Furfural	3	6	-	-
Solubles	6	20	1	-
Insolubles	2,124	106	2,124	-
Sulfuric Acid	-	-	-	-
Sodium Hydroxide	-	-	-	-
Sodium Sulfate	106	383	16	-
Enzyme	738	737	111	-
Mycelia	223	12	223	-
Nutrients	37	74	6	-
Ammonium Hydroxide	52	103	8	-
Carbon Dioxide	-	-	-	-
Air	-	-	-	-
Total	187,477	424,531	38,071	36,714
Temperature, °F	122	180	122	80

TABLE 5.1-1 (Cont)

(Section 400 - Hydrolysis)

Stream Number	413	414	415	416
Stream Name	Solids to Boiler	Water Wash Product to Evaporation	Hydrolysis Recycle Centrifuge Effluent to Evaporation	pH Adjustment of Recovery Tank
	lb/hr	lb/hr	lb/hr	lb/hr
Component				
Water	24,803	36,406	359,367	53
Cellulose	3,940	207	-	-
Glucose	91	1,522	22,579	-
Other Hexose	1	8	201	-
Pentose	10	96	2,443	-
Lignin	3,758	198	-	-
Ethanol	-	-	-	-
Degraded Pentose	1,193	63	-	-
Furfural	-	-	6	-
Solubles	-	1	19	-
Insolubles	2,018	106	-	-
Sulfuric Acid	-	-	-	123
Sodium Hydroxide	-	-	-	-
Sodium Sulfate	2	14	369	-
Enzyme	6	105	632	-
Mycelia	211	12	-	-
Nutrients	1	5	69	-
Ammonium Hydroxide	1	7	96	-
Carbon Dioxide	-	-	-	-
Air	-	-	-	-
Total	36,035	38,750	385,781	176
Temperature, °F	101	101	120	80

TABLE 5.1-1 (Cont)

(Section 400 - Hydrolysis)

Stream Number	417	418	419	420
Stream Name	pH Adjustment to Hydrolysis	Water to Hydrolysis	Makeup Water to Hydrolysis	Pretreated Cellulose Feed to Enzyme Production
Component	lb/hr	lb/hr	lb/hr	lb/hr
Water	-	265,000	33,522	9,248
Cellulose	-	-	-	2,226
Glucose	-	-	-	-
Other Hexose	-	-	-	17
Pentose	-	-	-	203
Lignin	-	-	-	315
Ethanol	-	-	-	-
Degraded Pentose	-	-	-	100
Furfural	-	-	-	-
Solubles	-	-	-	2
Insolubles	-	-	-	169
Sulfuric Acid	123	-	-	-
Sodium Hydroxide	-	-	-	17
Sodium Sulfate	-	-	-	-
Enzyme	-	-	-	-
Mycelia	-	-	-	-
Nutrients	-	-	-	-
Ammonium Hydroxide	-	-	-	-
Carbon Dioxide	-	-	-	-
Air	-	-	-	-
Total	123	265,000	33,522	12,297
Temperature, °F	80	131	80	94

Stream Number	421
Stream Name	Acid Dilution Water
Component	lb/hr
Water	53
Cellulose	-
Glucose	-
Other Hexose	-
Pentose	-
Lignin	-
Ethanol	-
Degraded Pentose	-
Furfural	-
Solubles	-
Insolubles	-
Sulfuric Acid	-
Sulfuric Hydroxide	-
Sodium Sulfate	-
Enzyme	-
Mycella	-
Nutrients	-
Ammonium Hydroxide	-
Carbon Dioxide	-
Air	-
Fusel Oils	-
Other Volatiles	-
Gasoline	-
Ash	-
Total	53
Temperature, °F	80

TABLE 5.1-1 (Cont)

(Section 500 - Evaporation)

Stream Number Stream Name	501 Evaporator Condensate to Chip Soak	502 Evaporator Condensate to Hydrolysis	503 Evaporator Product to Fermentation	504 LP Steam (65 psia)	505 Condensate
Component					
Water	27,464	231,531	136,778	58,780	58,780
Cellulose	-	-	207	-	-
Glucose	-	-	24,309	-	-
Other Hexose	-	-	combined as glucose	-	-
Pentose	-	-	2,539	-	-
Lignin	-	-	198	-	-
Ethanol	-	-	-	-	-
Degraded Pentose	-	-	63	-	-
Furfural	-	-	6	-	-
Solubles	-	-	20	-	-
Insolubles	-	-	106	-	-
Sulfuric Acid	-	-	-	-	-
Sodium Hydroxide	-	-	-	-	-
Sodium Sulfate	-	-	383	-	-
Enzyme	-	-	737	-	-
Mycelia	-	-	-	-	-
Nutrients	-	-	74	-	-
Ammonium Hydroxide	-	-	103	-	-
Carbon Dioxide	-	-	-	-	-
Air	-	-	-	-	-
Fusel Oils	-	-	-	-	-
Other Volatiles	-	-	-	-	-
Gasoline	-	-	-	-	-
Ash	-	-	-	-	-
Total	27,464	231,531	165,523	58,780	58,780
Temperature, °F	220	166	170	-	298

TABLE 5.1-1 (Cont)

(Section 600 - Fermentation)

Stream Number	601	602	603	604
Stream Name	Fermenter Feed	Cooled Fermenter Feed	CO_2 from Fermenter	Fermenter Broth to Beer Still
Component	lb/hr	lb/hr	lb/hr	lb/hr
Water	136,778	136,778	185	136,593
Cellulose	207	207	-	207
Glucose	24,309	24,309	-	486
Other Hexose	-	-	-	729
Pentose	2,539	2,539	-	2,539
Lignin	198	198	-	198
Ethanol	-	-	46	11,758
Degraded Pentose	63	63	-	65
Furfural	6	6	-	6
Solubles	20	20	-	20
Insolubles	106	106	-	106
Sulfuric Acid	-	-	-	-
Sodium Hydroxide	-	-	-	-
Sodium Sulfate	383	383	-	383
Enzyme	737	737	-	737
Mycelia	-	-	-	-
Nutrients	74	74	-	74
Ammonium Hydroxide	103	103	-	103
Carbon Dioxide	-	-	11,120	170
Air	-	-	-	-
Fusel Oils	-	-	-	35
Other Volatiles	-	-	-	16
Gasoline	-	-	-	-
Ash	-	-	-	-
Total	165,523	165,523	11,351	154,225
Temperature, °F	105	85	85	85

TABLE 5.1-1 (Cont)

(Section 700 - Distillation)

Stream Number	701	702	703	704	705
Stream Name	Beer Still Bottoms	Beer Still Bottoms to Vacuum Flash	Ethanol Product	Waste Water	Beer Still Recycle to Wash
Component	lb/hr	lb/hr	lb/hr	lb/hr	lb/hr
Water	167,572	32,059	96	925	135,513
Cellulose	207	40	-	-	167
Glucose	486	93	-	-	393
Other Hexose	729	139	-	-	590
Pentose	2,539	486	-	-	2,053
Lignin	198	38	-	-	160
Ethanol	-	-	11,775	6	-
Degraded Pentose	65	12	-	-	53
Furfural	6	1	-	-	5
Solubles	20	4	-	-	16
Insolubles	106	20	-	-	86
Sulfuric Acid	-	-	-	-	-
Sodium Hydroxide	-	-	-	-	-
Sodium Sulfate	383	73	-	-	310
Enzyme	737	141	-	-	596
Mycella	-	-	-	-	-
Nutrients	74	14	-	-	60
Ammonium Hydroxide	103	20	-	-	83
Carbon Dioxide	-	-	-	-	-
Air	-	-	-	-	-
Fusel Oils	-	-	-	-	-
Other Volatiles	-	-	-	7	-
Gasoline	-	-	-	-	-
Ash	-	-	-	-	-
Total	173,225	33,140	11,871	938	140,085
Temperature, °F	239	90	100	229	90

TABLE 5.1-1 (Cont)

(Section 700 - Distillation)

Stream Number	706	707	708
Stream Name	Fusel Oils to Product Blending	Vents to Vent Recovery	Azeotrope to Anhydrous Column
Component	lb/hr	lb/hr	lb/hr
Water	6	2	1,021
Cellulose	-	-	-
Glucose	-	-	-
Other Hexose	-	-	-
Pentose	-	-	-
Lignin	-	-	-
Ethanol	17	35	11,781
Degraded Pentose	-	-	-
Furfural	-	-	-
Solubles	-	-	-
Insolubles	-	-	-
Sulfuric Acid	-	-	-
Sodium Hydroxide	-	-	-
Sodium Sulfate	-	-	-
Enzyme	-	-	-
Mycelia	-	-	-
Nutrients	-	-	-
Ammonium Hydroxide	-	-	-
Carbon Dioxide	-	170	-
Air	-	-	-
Fusel Oils	35	-	-
Other Volatiles	7	19	7
Gasoline	-	-	-
Ash	-	-	-
Total	65	226	12,809
Temperature, °F	88	85	174

TABLE 5.1-1 (Cont)

(Section 800 - Anaerobic Digestion)

Stream Number Stream Name Component	801 Feed to Digester lb/hr	804 Biogas lb/hr	805 Waste lb/hr
Water	161,860	-	161,860
Cellulose	207	-	-
Glucose	486	-	-
Other Hexose	1,248	-	-
Pentose	8,869	-	-
Lignin	1,212	-	-
Ethanol	-	-	-
Degraded Pentose	65	-	COD 2,088
Furfural	238	-	-
Solubles	938	-	-
Insolubles	107	-	-
Sulfuric Acid	-	-	-
Sodium Hydroxide	-	-	-
Sodium Sulfate	385	-	-
Enzyme	737	-	-
Mycelia	-	-	-
Nutrients	74	-	-
Ammonium Hydroxide	8	-	-
Carbon Dioxide	-	5,138	-
Air	-	-	-
Methane	-	2,803	-
Sludge	-	-	-
Total	176,434	7,941	-
Temperature, °F	105	98	98

(Section 900 - Boiler)

Stream Number	902	903	904	905	906
Stream Name	Feed to Lignin Centrifuge	Feed to Lignin Boiler	Waste from Lignin Centrifuge	Total Lignin Feed to Boiler	HP Steam (665 psia)
Component	lb/hr	lb/hr	lb/hr	lb/hr	lb/hr
Water	121,494	11,386	110,108	36,189	131,608
Cellulose	-	-	-	3,940	-
Glucose	-	-	-	91	-
Other Hexose	35	3	32	4	-
Pentose	430	40	390	50	-
Lignin	6,957	6,610	347	10,368	-
Ethanol	-	-	-	-	-
Degraded Pentose	-	-	-	1,193	-
Furfural	35	35	-	35	-
Solubles	111	10	101	10	-
Insolubles	-	-	-	2,018	-
Sulfuric Acid	-	-	-	-	-
Sodium Hydroxide	-	-	-	-	-
Sodium Sulfate	2,180	203	1,977	205	-
Enzyme	-	-	-	6	-
Mycelia	-	-	-	211	-
Nutrients	-	-	-	1	-
Ammonium Hydroxide	-	-	-	1	-
Carbon Dioxide	-	-	-	-	-
Air	-	-	-	-	-
Fusel Oils	-	-	-	-	-
Other Volatiles	-	-	-	-	-
Gasoline	-	-	-	-	-
Ash	-	-	-	-	-
Total	131,242	18,287	112,955	54,322	131,608
Temperature, °F	101	101	101	101	750

TABLE 5.1-1 (Cont)

(Section 900 - Boiler)

Stream Number	908	909
Stream Name	Ash	Wood Chips to Boiler
	lb/hr	lb/hr
Component		
Water	-	3,475
Cellulose	-	1,783
Glucose	-	-
Other Hexose	-	49
Pentose	-	691
Lignin	-	761
Ethanol	-	-
Degraded Pentose	-	-
Furfural	-	-
Solubles	-	56
Insolubles	-	135
Sulfuric Acid	-	-
Sodium Hydroxide	-	-
Sodium Sulfate	-	-
Enzyme	-	-
Mycelia	-	-
Nutrients	-	-
Ammonium Hydroxide	-	-
Carbon Dioxide	-	-
Air	-	-
Fusel Oils	-	-
Other Volatiles	-	-
Gasoline	-	-
Ash	2,448	-
Total	2,448	6,950
Temperature, °F	-	80

TABLE 5.1-1 (Cont)

(Section 1000 - Feedstock Handling)

Stream Number	1001	1002
Stream Name	Wood Chip Feed	Wood Chips to Boiler
Component		
Water	54,462	3,475
Cellulose	27,939	1,783
Glucose	-	-
Hexose	763	49
Pentose	10,838	691
Lignin	11,927	761
Ethanol	-	-
Degraded Pentose	-	-
Furfural	-	-
Solubles	871	56
Insolubles	2,124	135
Sulfuric Acid	-	-
Sodium Hydroxide	-	-
Sodium Sulfate	-	-
Enzyme	-	-
Mycelia	-	-
Nutrients	-	-
Ammonium Hydroxide	-	-
Carbon Dioxide	-	-
Air	-	-
Fusel Oils	-	-
Other Volatiles	-	-
Gasoline	-	-
Ash	-	-
Total	108,924	6,950
Temperature, °F	80	80

TABLE 5.1-1 (Cont)

(Section 1200 - Waste Treatment/Vent Scrubbing)

Stream Number	1201	1202	1203	1204	1205	1206	1207
Stream Name	Sludge for Disposal	Treated Water Discharge	Vent Column Wash Water	Vent Column Vent	Recovered Vents	CO_2 Wash Column Wash Water	CO_2 Column Vent
Component							
Water	-	167,224	745	-	745	5,585	185
Cellulose	-	-	-	-	-	-	-
Glucose	-	-	-	-	-	-	-
Hexose	-	-	-	-	-	-	-
Pentose	-	-	-	-	-	-	-
Lignin	-	-	-	-	-	-	-
Ethanol	-	-	-	4	31	-	2
Degraded Pentose	-	-	-	-	-	-	-
Furfural	-	-	-	-	-	-	-
Solubles	-	-	-	-	-	-	-
Insolubles	1,238 (sludge)	-	-	-	-	-	-
Sulfuric Acid	-	-	-	-	-	-	-
Sodium Hydroxide	-	-	-	-	-	-	-
Sodium Sulfate	-	-	-	-	-	-	-
Enzyme	-	-	-	-	-	-	-
Mycelia	-	-	-	-	-	-	-
Nutrients	-	-	-	-	-	-	-
Ammonium Hydroxide	-	-	-	-	-	-	-
Carbon Dioxide	-	-	-	170	-	-	11,120
Air	-	-	-	-	-	-	-
Fusel Oils	-	-	-	-	-	-	-
Other Volatiles	-	-	-	2	17	-	-
Gasoline	-	-	-	-	-	-	-
Ash	-	-	-	-	-	-	-
Total	1,238	167,224	745	176	793	5,585	11,307
Temperature, °F	80	80	80	80	80	80	80

TABLE 5.1-1 (Cont)

(Section 1200 - Waste Treatment/Vent Scrubbing)

Stream Number	1208
Stream Name	Recovered CO_2 Vents
Component	
Water	5,585
Cellulose	-
Glucose	-
Hexose	-
Pentose	-
Lignin	-
Ethanol	44
Degraded Pentose	-
Furfural	-
Solubles	-
Insolubles	-
Sulfuric Acid	-
Sodium Hydroxide	-
Sodium Sulfate	-
Enzyme	-
Mycelia	-
Nutrients	-
Ammonium Hydroxide	-
Carbon Dioxide	-
Air	-
Fusel Oils	-
Other Volatiles	-
Gasoline	-
Ash	-
Total	5,629
Temperature, °F	80

TABLE 5.1-1 (Cont)

(Section 1300 - Chemical Handling)

Stream Number	1301	1302	1303	1304	1305	1306
Stream Name	NaOH to Water/ Alkali Wash	H_2SO_4 to Pretreatment	H_2SO_4 to Enzyme Fermenters	H_2SO_4 to Hydrolysis	H_2SO_4 to Boiler	H_2SO_4 to Plant
Component						
Water	1,443	-	-	-	-	-
Cellulose	-	-	-	-	-	-
Glucose	-	-	-	-	-	-
Hexose	-	-	-	-	-	-
Pentose	-	-	-	-	-	-
Lignin	-	-	-	-	-	-
Ethanol	-	-	-	-	-	-
Degraded Pentose	-	-	-	-	-	-
Furfural	-	-	-	-	-	-
Solubles	-	-	-	-	-	-
Insolubles	-	-	-	-	-	-
Sulfuric Acid	1,443	156	21	246	1,485	1,908
Sodium Hydroxide	-	-	-	-	-	-
Sodium Sulfate	-	-	-	-	-	-
Enzyme	-	-	-	-	-	-
Mycelia	-	-	-	-	-	-
Nutrients	-	-	-	-	-	-
Ammonium Hydroxide	-	-	-	-	-	-
Carbon Dioxide	-	-	-	-	-	-
Air	-	-	-	-	-	-
Fusel Oils	-	-	-	-	-	-
Other Volatiles	-	-	-	-	-	-
Gasoline	-	-	-	-	-	-
Ash	-	-	-	-	-	-
	-	-	-	-	-	-
Total	2,886	156	21	246	1,485	1,908
Temperature, °F	80	80	80	80	80	80

TABLE 5.1-1 (Cont)

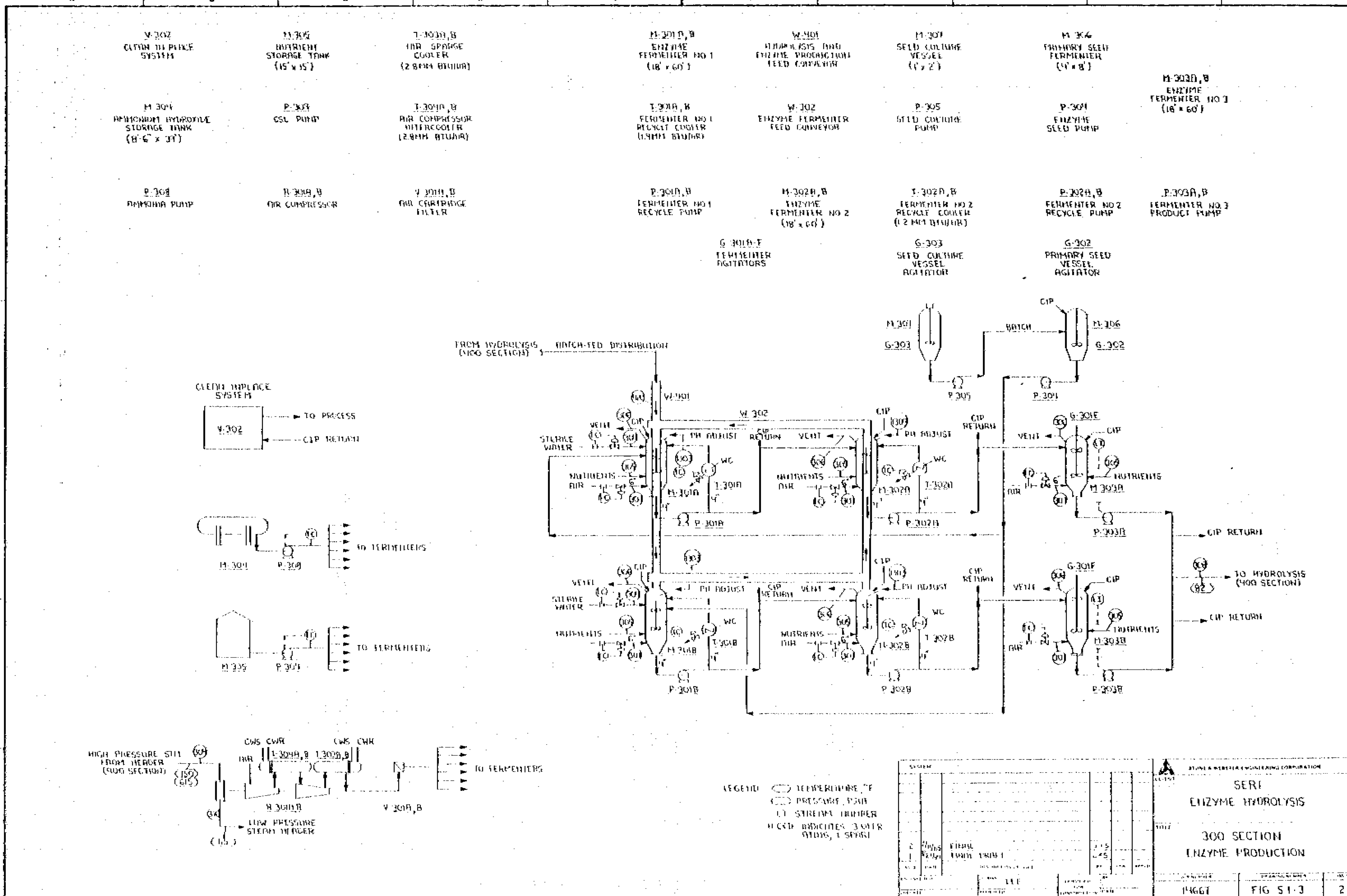
(Section 1400 - Product Storage and Loading)

Stream Number	1401	1402
Stream Name	Gasoline	Denatured Ethanol Product
Component		
Water	-	102
Cellulose	-	-
Glucose	-	-
Hexose	-	-
Pentose	-	-
Lignin	-	-
Ethanol	-	11,792
Degraded Pentose	-	-
Furfural	-	-
Solubles	-	-
Insolubles	-	-
Sulfuric Acid	-	-
Sodium Hydroxide	-	-
Sodium Sulfate	-	-
Enzyme	-	-
Mycelia	-	-
Nutrients	-	-
Ammonium Hydroxide	-	-
Carbon Dioxide	-	-
Air	-	-
Fusel Oils	-	35
Other Volatiles	-	7
Gasoline	613	613
Ash	-	-
Total	613	12,549
Temperature, °F	80	80

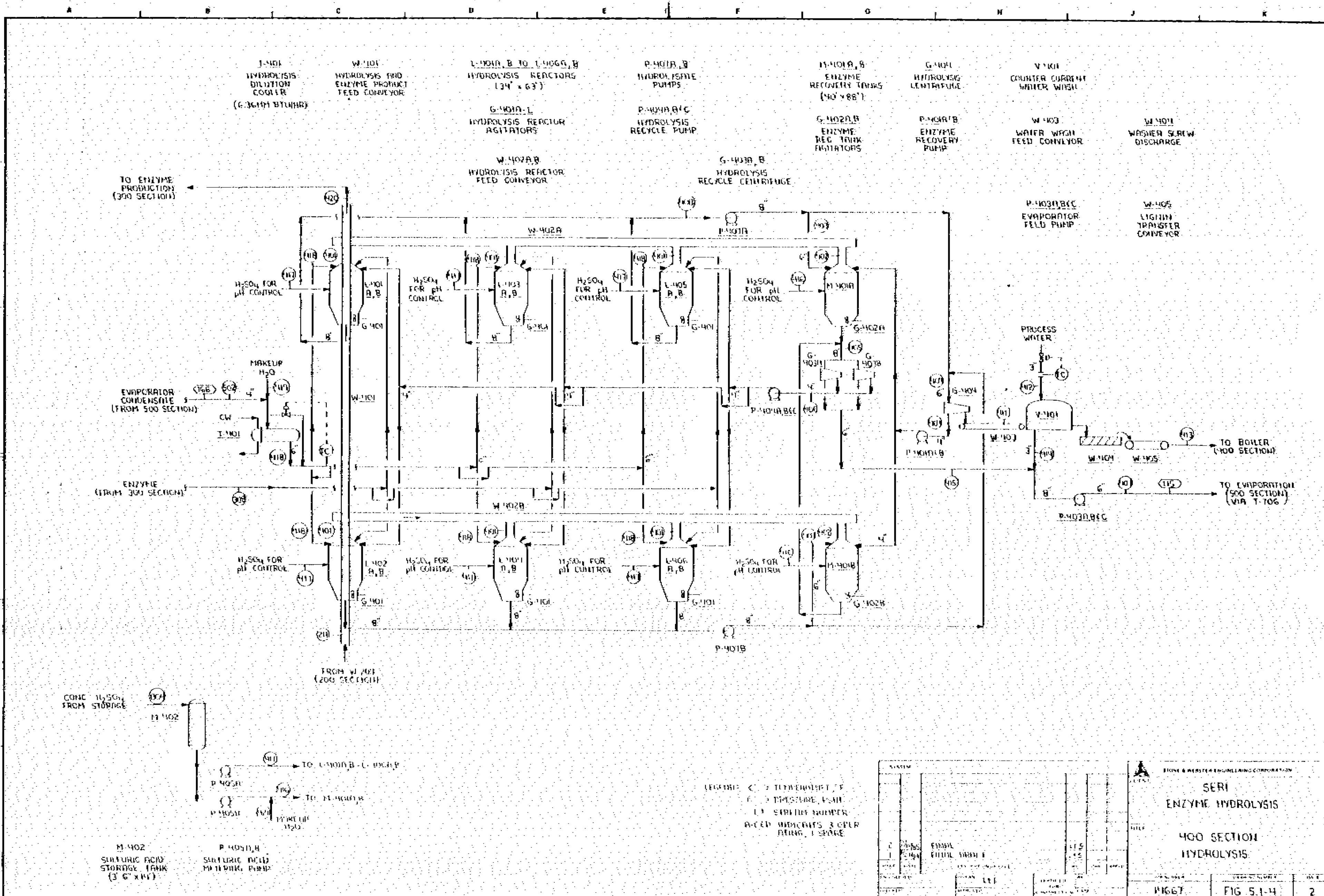
TABLE 3.1-1 (Cont)

(Section 1500 - Waste Treatment/Condensate Return)

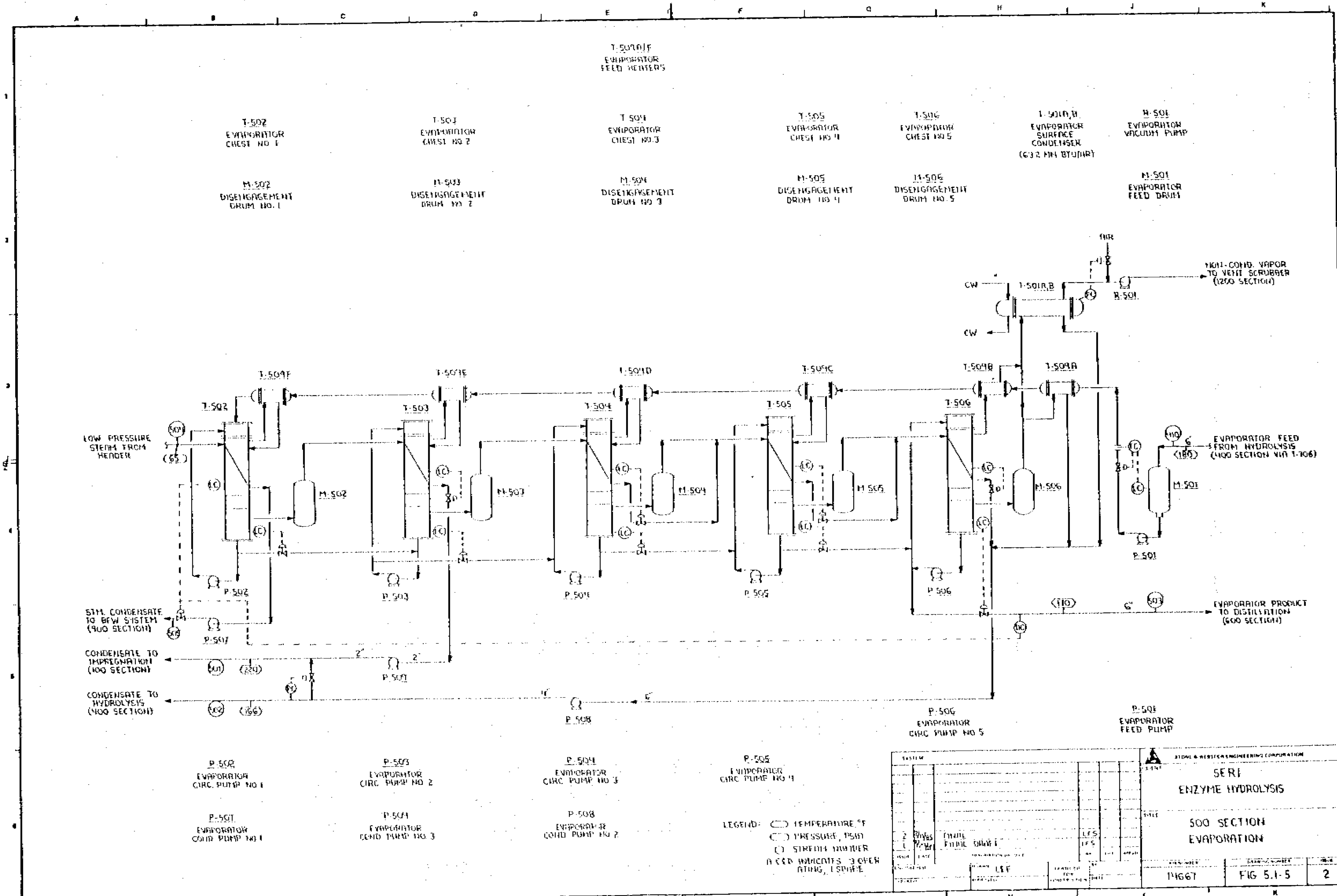
Stream Number	1501	1502
Stream Name	Boiler Feed Water	Makeup Water
Component		
Water	131,608	49,070
Cellulose	-	-
Glucose	-	-
Hexose	-	-
Pentose	-	-
Lignin	-	-
Ethanol	-	-
Degraded Pentose	-	-
Furfural	-	-
Solubles	-	-
Insolubles	-	-
Sulfuric Acid	-	-
Sodium Hydroxide	-	-
Sodium Sulfate	-	-
Enzyme	-	-
Mycelia	-	-
Nutrients	-	-
Ammonium Hydroxide	-	-
Carbon Dioxide	-	-
Air	-	-
Fusel Oils	-	-
Other Volatiles	-	-
Gasoline	-	-
Ash	-	-
Total	131,608	49,070
Temperature, °F	250	80



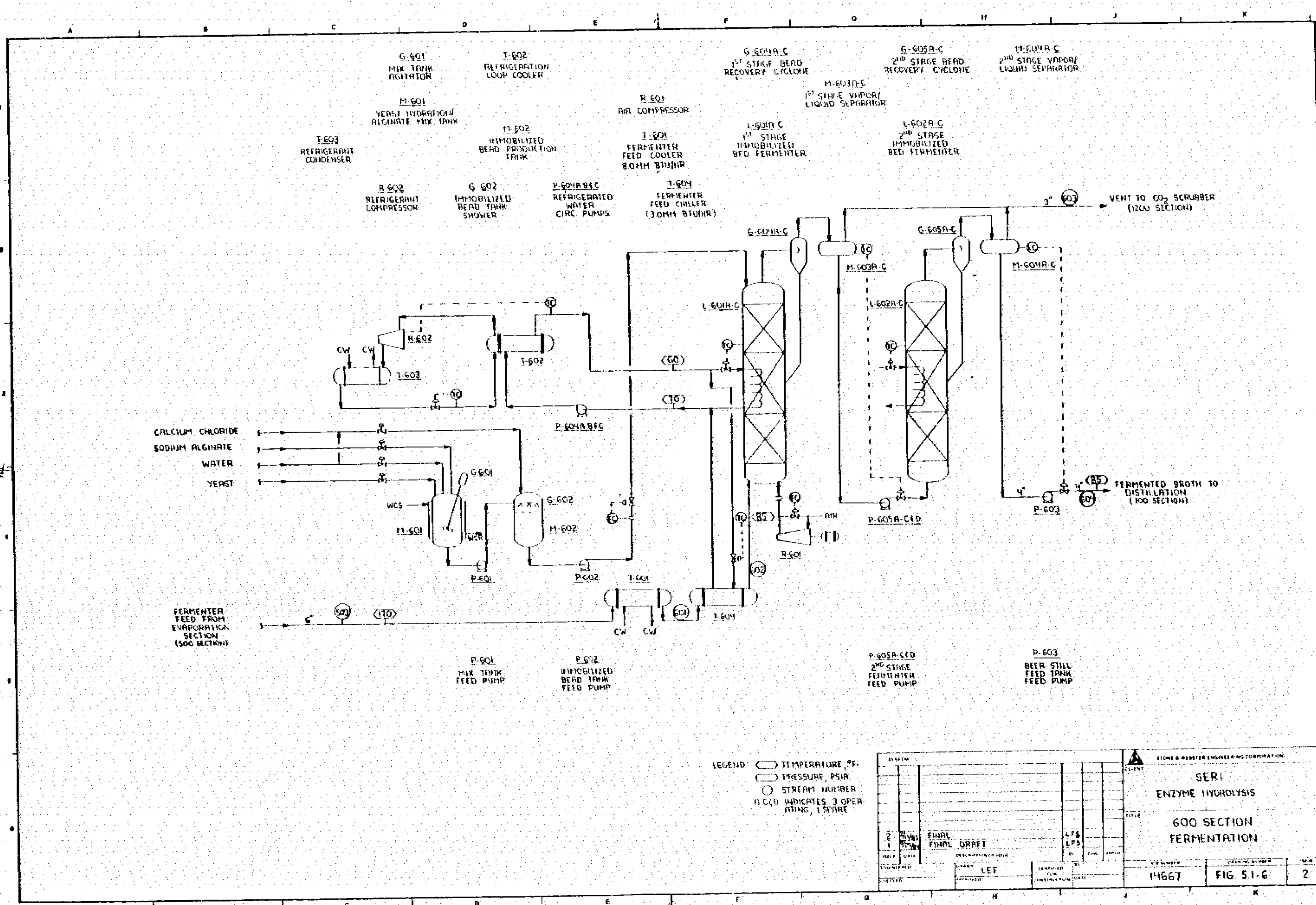
V-302 CLEAN IN PLACE SYSTEM
M-305 NUTRIENT STORAGE TANK (15' x 15')
T-303A,B AIR SPARGE COOLER
M-301A,B ENZYME FERMENTER NO 1 (18' x 60')
W-301 HYDROLYSIS AND ENZYME PRODUCTION FEED CONVEYOR
M-307 SEED CULTURE VESSEL
M-306 PRIMARY SEED FERMENTER
M-303A,B ENZYME FERMENTER NO 3 (18' x 60')
M-304 AMMONIUM HYDROXIDE STORAGE TANK
P-303 CSL PUMP
T-304A,B AIR COMPRESSOR INTERCOOLER
T-301A,B FERMENTER NO 1 RECYCLE COOLER
W-302 ENZYME FERMENTER FEED CONVEYOR
P-305 SEED CULTURE PUMP
P-304 ENZYME SEED PUMP
P-308 AMMONIA PUMP
B-301A,B AIR COMPRESSOR
V-301A,B AIR CARTRIDGE FILTER
P-301A,B FERMENTER NO 1 RECYCLE PUMP
M-302A,B ENZYME FERMENTER NO 2 (18' x 60')
T-302A,B FERMENTER NO 2 RECYCLE COOLER
P-302A,B FERMENTER NO 2 RECYCLE PUMP
P-303A,B FERMENTER NO 3 PRODUCT PUMP
G-301A-F FERMENTER AGITATORS
G-303 SEED CULTURE VESSEL AGITATOR
G-302 PRIMARY SEED VESSEL AGITATOR
CLEAN INPLACE SYSTEM
TO PROCESS
CIP RETURN
TO FERMENTERS
FROM HYDROLYSIS (400 SECTION)
BATCH-FED DISTRIBUTION
BATCH
STERILE WATER
NUTRIENTS
AIR
VENT
CIP
PH ADJUST
CIP RETURN
TO HYDROLYSIS (400 SECTION)
HIGH PRESSURE STM FROM HEADER (400 SECTION)
LOW PRESSURE STEAM HEADER
CWS CWR
LEGEND
TEMPERATURE °F
PRESSURE, PSIA
STREAM NUMBER
SERI
ENZYME HYDROLYSIS
300 SECTION
ENZYME PRODUCTION
FIG S1-3

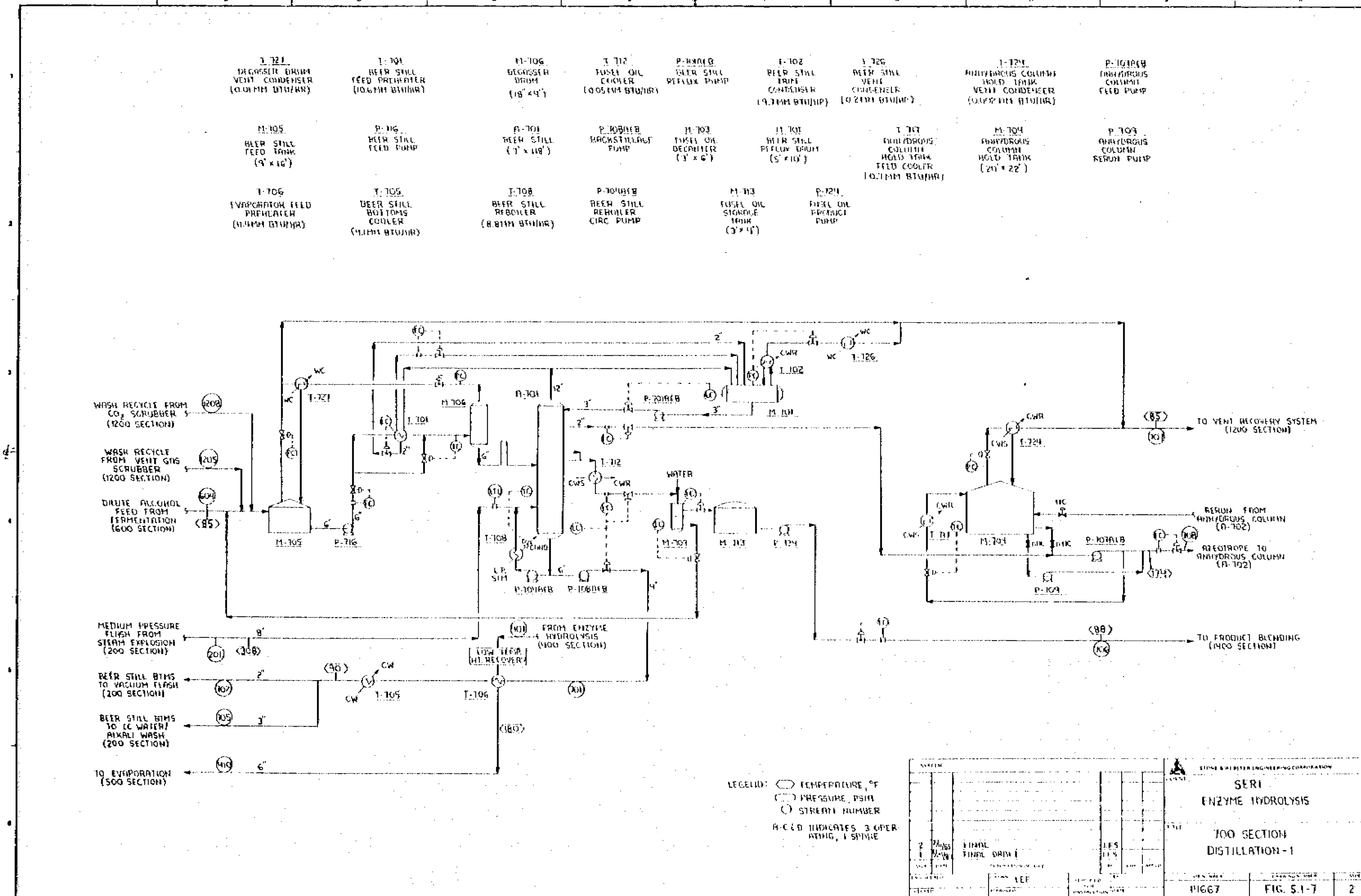


T-401 HYDROLYSIS DILUTION COOLER (6.36MM BTU/HR)
W-401 HYDROLYSIS AND ENZYME PRODUCT FEED CONVEYOR
L-401A,B TO L-406A,B HYDROLYSIS REACTORS (34' x 63')
G-401A-L HYDROLYSIS REACTOR AGITATORS
W-402A,B HYDROLYSIS REACTOR FEED CONVEYOR
P-401A,B HYDROLYSATE PUMPS
P-404A,B,C HYDROLYSIS RECYCLE PUMP
G-403A,B HYDROLYSIS RECYCLE CENTRIFUGE
M-401A,B ENZYME RECOVERY TANKS (40' x 86')
G-402A,B ENZYME REC TANK AGITATORS
G-404 HYDROLYSIS CENTRIFUGE
P-402A/B ENZYME RECOVERY PUMP
V-401 COUNTER CURRENT WATER WASH
W-403 WATER WASH FEED CONVEYOR
W-404 WASHER SCREW DISCHARGE
P-403A,B,C EVAPORATOR FEED PUMP
W-405 LIGNIN TRANSFER CONVEYOR
TO ENZYME PRODUCTION (300 SECTION)
EVAPORATOR CONDENSATE (FROM 500 SECTION)
MAKEUP H_2O
CW
ENZYME (FROM 300 SECTION)
H_2SO_4 FOR pH CONTROL
W-402A
W-402B
W-401
PROCESS WATER
TO BOILER (400 SECTION)
TO EVAPORATION (500 SECTION) VIA T-706
FROM W-203 (200 SECTION)
P-401B
CONC H_2SO_4 FROM STORAGE
M-402
P-405A
P-405B
TO L-401A,B - L-406A,B
TO M-401A,B
MAKEUP H_2O
M-402 SULFURIC ACID STORAGE TANK (3' 6" x 14')
P-405A,B SULFURIC ACID METERING PUMP
LEGEND: ⟨ ⟩ TEMPERATURE, °F
( ) PRESSURE, PSIA
( ) STREAM NUMBER
A-C,D INDICATES 3 OPERATING, 1 SPARE
STONE & WEBSTER ENGINEERING CORPORATION
SERI
ENZYME HYDROLYSIS
400 SECTION
HYDROLYSIS
FINAL
FINAL DRAFT
P1667
FIG 5.1-4
2

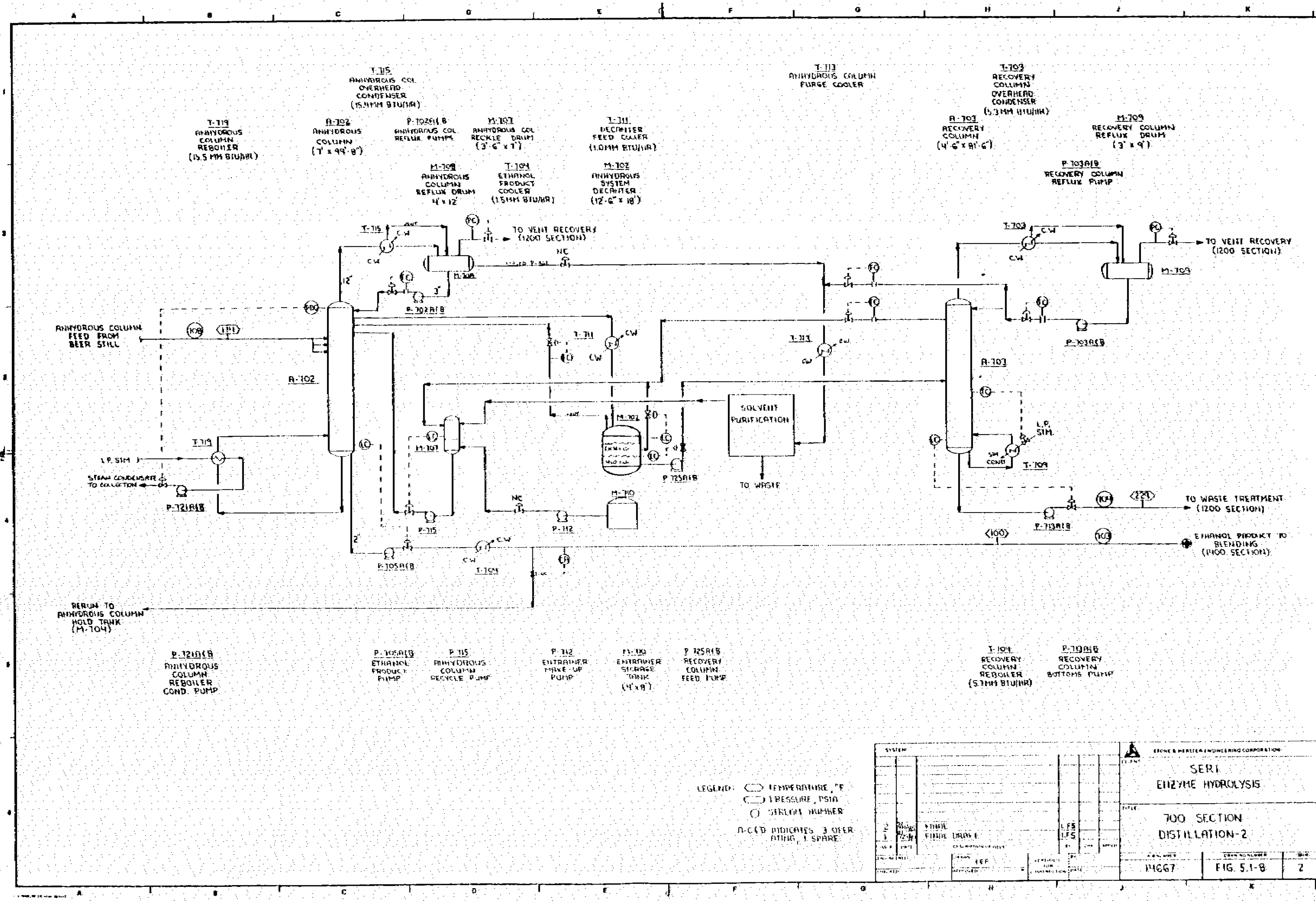


T-509A-F EVAPORATOR FEED HEATERS
T-502 EVAPORATOR CHEST NO 1
T-503 EVAPORATOR CHEST NO 2
T-504 EVAPORATOR CHEST NO 3
T-505 EVAPORATOR CHEST NO 4
T-506 EVAPORATOR CHEST NO 5
T-501A,B EVAPORATOR SURFACE CONDENSER (63.2 MM BTU/HR)
R-501 EVAPORATOR VACUUM PUMP
M-502 DISENGAGEMENT DRUM NO 1
M-503 DISENGAGEMENT DRUM NO 2
M-504 DISENGAGEMENT DRUM NO 3
M-505 DISENGAGEMENT DRUM NO 4
M-506 DISENGAGEMENT DRUM NO 5
M-501 EVAPORATOR FEED DRUM
AIR
NON-COND. VAPOR TO VENT SCRUBBER (1200 SECTION)
CW
T-501A,B
R-501
T-509F
T-509E
T-509D
T-509C
T-509B
T-509A
T-502
T-503
T-504
T-505
T-506
LOW PRESSURE STEAM FROM HEADER
EVAPORATOR FEED FROM HYDROLYSIS (400 SECTION VIA T-406)
M-502
M-503
M-504
M-505
M-506
M-501
P-501
P-502
P-503
P-504
P-505
P-506
P-507
P-509
P-508
STM. CONDENSATE TO BFW SYSTEM (900 SECTION)
CONDENSATE TO IMPREGNATION (100 SECTION)
CONDENSATE TO HYDROLYSIS (400 SECTION)
EVAPORATOR PRODUCT TO DISTILLATION (600 SECTION)
P-506 EVAPORATOR CIRC PUMP NO 5
P-501 EVAPORATOR FEED PUMP
P-502 EVAPORATOR CIRC. PUMP NO 1
P-503 EVAPORATOR CIRC. PUMP NO 2
P-504 EVAPORATOR CIRC. PUMP NO 3
P-505 EVAPORATOR CIRC. PUMP NO 4
P-507 EVAPORATOR COND PUMP NO 1
P-509 EVAPORATOR COND PUMP NO 3
P-508 EVAPORATOR COND PUMP NO 2
LEGEND: ⬭ TEMPERATURE, °F
⬭ PRESSURE, PSIA
○ STREAM NUMBER
A (2) INDICATES 3 OPERATING, 1 SPARE
STONE & WEBSTER ENGINEERING CORPORATION
SERI
ENZYME HYDROLYSIS
500 SECTION
EVAPORATION
14667
FIG 5.1-5
2

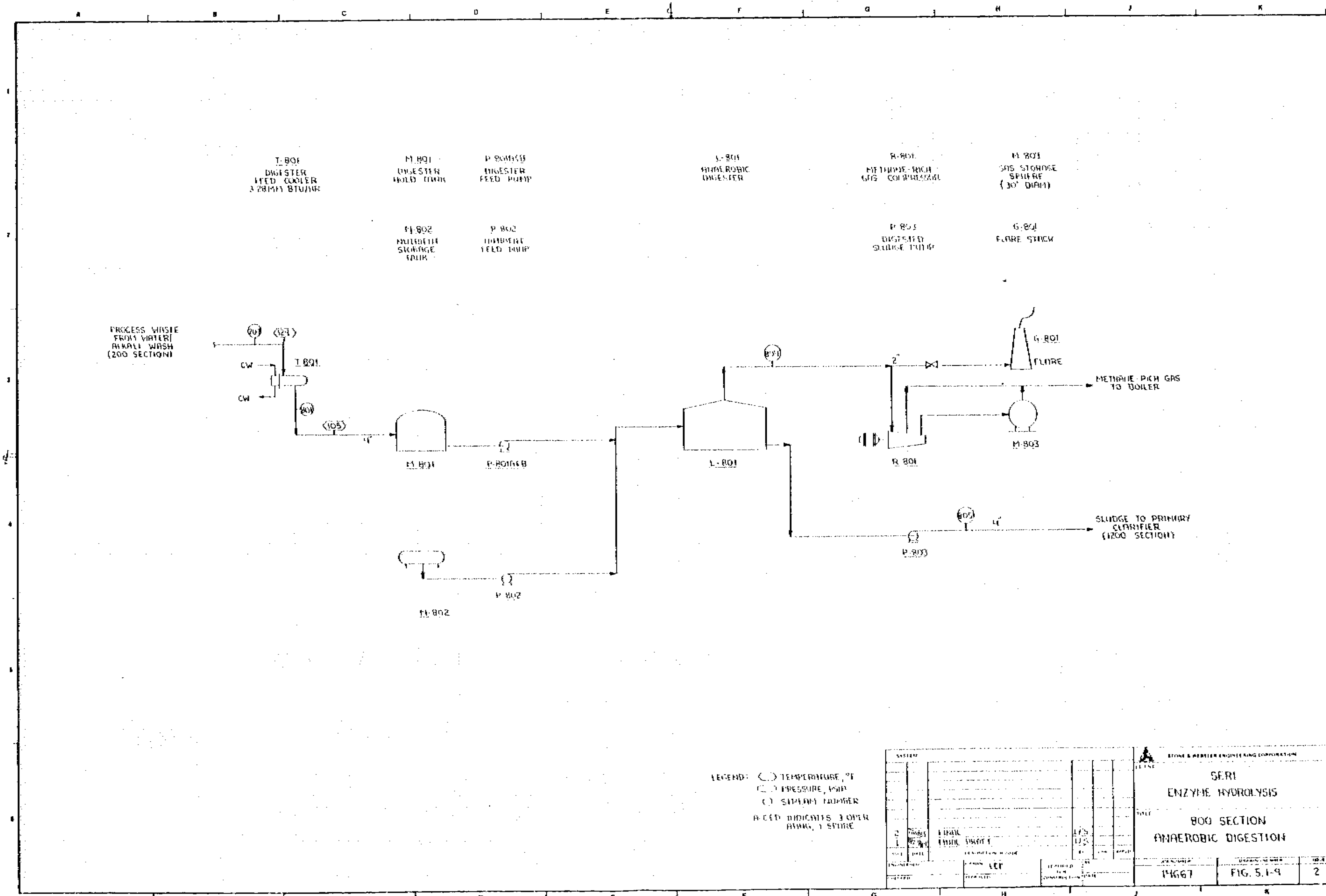

G-601 MIX TANK AGITATOR
T-602 REFRIGERATION LOOP COOLER
M-601 YEAST HYDRATION/ ALGINATE MIX TANK
M-602 IMMOBILIZED BEAD PRODUCTION TANK
T-603 REFRIGERANT CONDENSER
R-602 REFRIGERANT COMPRESSOR
G-602 IMMOBILIZED BEAD TANK SHOWER
P-604A,B&C REFRIGERATED WATER CIRC. PUMPS
R-601 AIR COMPRESSOR
T-601 FERMENTER FEED COOLER 80MM BTU/HR
T-604 FERMENTER FEED CHILLER (30MM BTU/HR)
G-604A-C 1ST STAGE BEAD RECOVERY CYCLONE
M-603A-C 1ST STAGE VAPOR/ LIQUID SEPARATOR
L-601A-C 1ST STAGE IMMOBILIZED BED FERMENTER
G-605A-C 2ND STAGE BEAD RECOVERY CYCLONE
L-602A-C 2ND STAGE IMMOBILIZED BED FERMENTER
M-604A-C 2ND STAGE VAPOR/ LIQUID SEPARATOR
VENT TO CO_2 SCRUBBER (1200 SECTION)
CALCIUM CHLORIDE
SODIUM ALGINATE
WATER
YEAST
FERMENTER FEED FROM EVAPORATION SECTION (500 SECTION)
FERMENTED BROTH TO DISTILLATION (100 SECTION)
AIR
CW
P-601 MIX TANK FEED PUMP
P-602 IMMOBILIZED BEAD TANK FEED PUMP
P-605A-C&D 2ND STAGE FERMENTER FEED PUMP
P-603 BEER STILL FEED TANK FEED PUMP
LEGEND: TEMPERATURE, °F. PRESSURE, PSIA STREAM NUMBER
A,C&D INDICATES 3 OPERATING, 1 SPARE
STONE & WEBSTER ENGINEERING CORPORATION
SERI
ENZYME HYDROLYSIS
600 SECTION
FERMENTATION
FINAL
FINAL DRAFT
LEF
14667
FIG. 5.1-6
2


T-721 DEGASSER DRUM VENT CONDENSER (0.01MM BTU/HR)
T-701 BEER STILL FEED PREHEATER (10.6MM BTU/HR)
M-706 DEGASSER DRUM (18" x 4')
T-712 FUSEL OIL COOLER (0.05MM BTU/HR)
BEER STILL REFLUX PUMP
T-702 BEER STILL TRIM CONDENSER (9.7MM BTU/HR)
T-726 BEER STILL VENT CONDENSER (0.21MM BTU/HR)
T-724 ANHYDROUS COLUMN HOLD TANK VENT CONDENSER (0.002 MM BTU/HR)
P-101A&B ANHYDROUS COLUMN FEED PUMP
M-705 BEER STILL FEED TANK (9' x 16')
P-716 BEER STILL FEED PUMP
A-701 BEER STILL (7' x 118')
P-108A&B BACKSTILLAGE PUMP
M-703 FUSEL OIL DECANTER (3' x 6')
M-701 BEER STILL REFLUX DRUM (5' x 10')
ANHYDROUS COLUMN HOLD TANK FEED COOLER (0.7MM BTU/HR)
M-704 ANHYDROUS COLUMN HOLD TANK (20' x 22')
P-703 ANHYDROUS COLUMN RERUN PUMP
T-706 EVAPORATOR FEED PREHEATER (11.4MM BTU/HR)
T-705 BEER STILL BOTTOMS COOLER (4.1MM BTU/HR)
T-708 BEER STILL REBOILER (8.81MM BTU/HR)
BEER STILL REBOILER CIRC PUMP
M-713 FUSEL OIL STORAGE TANK (3' x 4')
P-724 FUSEL OIL PRODUCT PUMP
WASH RECYCLE FROM CO2 SCRUBBER (1200 SECTION)
WASH RECYCLE FROM VENT GAS SCRUBBER (1200 SECTION)
DILUTE ALCOHOL FEED FROM FERMENTATION (600 SECTION)
MEDIUM PRESSURE FLASH FROM STEAM EXPLOSION (200 SECTION)
BEER STILL BTMS TO VACUUM FLASH (200 SECTION)
BEER STILL BTMS TO LC WATER/ALKALI WASH (200 SECTION)
TO EVAPORATION (500 SECTION)
FROM ENZYME HYDROLYSIS (100 SECTION)
LOW TEMP HT RECOVERY
TO VENT RECOVERY SYSTEM (1200 SECTION)
RERUN FROM ANHYDROUS COLUMN (A-702)
AZEOTROPE TO ANHYDROUS COLUMN (A-702)
TO PRODUCT BLENDING (1400 SECTION)
WATER
LEGEND: TEMPERATURE, °F
PRESSURE, PSIA
STREAM NUMBER
A-C&D INDICATES 3 OPERATING, 1 SPARE
STONE & WEBSTER ENGINEERING CORPORATION
SERI
ENZYME HYDROLYSIS
700 SECTION
DISTILLATION-1
FIG. 5.1-7
2

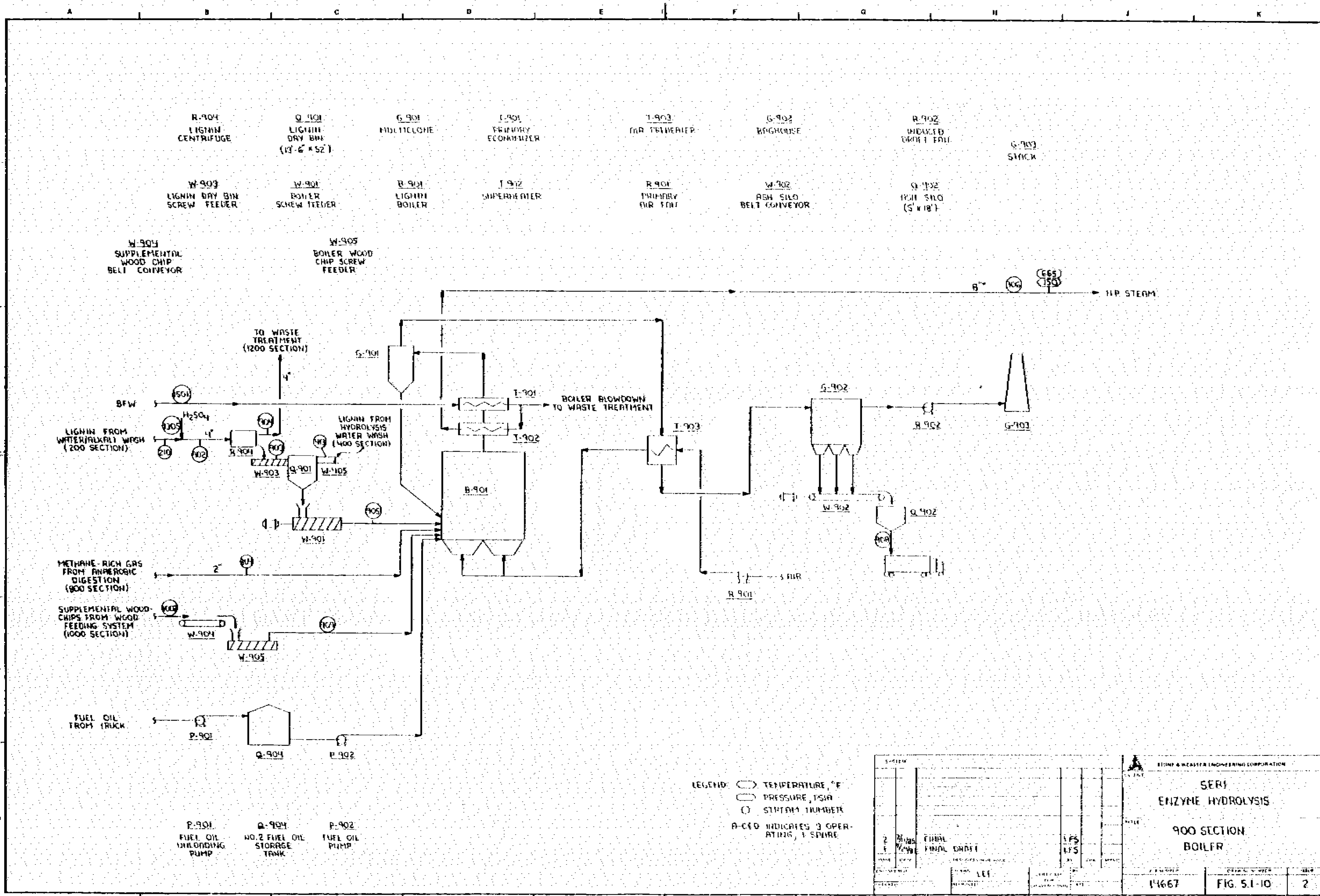
1



T-115 ANHYDROUS COL. OVERHEAD CONDENSER (15.1MM BTU/HR)
T-113 ANHYDROUS COLUMN PURGE COOLER
T-703 RECOVERY COLUMN OVERHEAD CONDENSER (5.3MM BTU/HR)
T-119 ANHYDROUS COLUMN REBOILER (15.5 MM BTU/HR)
A-702 ANHYDROUS COLUMN (7' x 99'-8")
P-702A&B ANHYDROUS COL. REFLUX PUMPS
M-707 ANHYDROUS COL. RECYCLE DRUM (3'-6" x 7')
T-711 DECANTER FEED COOLER (1.0MM BTU/HR)
A-703 RECOVERY COLUMN (4'-6" x 81'-6")
M-709 RECOVERY COLUMN REFLUX DRUM (3' x 9')
M-708 ANHYDROUS COLUMN REFLUX DRUM 4' x 12'
T-704 ETHANOL PRODUCT COOLER (1.5MM BTU/HR)
M-702 ANHYDROUS SYSTEM DECANTER (12'-6" x 18')
P-703A&B RECOVERY COLUMN REFLUX PUMP
TO VENT RECOVERY (1200 SECTION)
TO VENT RECOVERY (1200 SECTION)
ANHYDROUS COLUMN FEED FROM BEER STILL
SOLVENT PURIFICATION
L.P. STM
STEAM CONDENSATE TO COLLECTION
TO WASTE
TO WASTE TREATMENT (1200 SECTION)
ETHANOL PRODUCT TO BLENDING (1100 SECTION)
RERUN TO ANHYDROUS COLUMN HOLD TANK (M-704)
P-721A&B ANHYDROUS COLUMN REBOILER COND. PUMP
P-705A&B ETHANOL PRODUCT PUMP
P-715 ANHYDROUS COLUMN RECYCLE PUMP
P-712 ENTRAINER MAKE-UP PUMP
M-710 ENTRAINER STORAGE TANK (4' x 8')
P-725A&B RECOVERY COLUMN FEED PUMP
T-709 RECOVERY COLUMN REBOILER (5.7MM BTU/HR)
P-713A&B RECOVERY COLUMN BOTTOMS PUMP
LEGEND: TEMPERATURE, °F
PRESSURE, PSIA
STREAM NUMBER
A-C&D INDICATES 3 OPERATING, 1 SPARE
FINAL
FINAL DRAFT
SERI
ENZYME HYDROLYSIS
700 SECTION
DISTILLATION-2
14667
FIG. 5.1-8
2



T-801
DIGESTER FEED COOLER 3.28MM BTU/HR
M-801
DIGESTER HOLD TANK
P-801A&B
DIGESTER FEED PUMP
L-801
ANAEROBIC DIGESTER
R-801
METHANE-RICH GAS COMPRESSOR
M-803
GAS STORAGE SPHERE (30' DIAM)
M-802
NUTRIENT STORAGE TANK
P-802
NUTRIENT FEED PUMP
P-803
DIGESTED SLUDGE PUMP
G-801
FLARE STACK
PROCESS WASTE FROM WATER/ ALKALI WASH (200 SECTION)
CW
CW
T-801
(127)
(105)
4"
M-801
P-801A&B
M-802
P-802
L-801
2"
G-801
FLARE
METHANE-RICH GAS TO BOILER
R-801
M-803
4"
SLUDGE TO PRIMARY CLARIFIER (1200 SECTION)
P-803
LEGEND: ( ) TEMPERATURE, °F
( ) PRESSURE, PSIA
( ) STREAM NUMBER
A&B INDICATES 1 OPER. 1 SPARE
FINAL
FINAL DRAFT
STONE & WEBSTER ENGINEERING CORPORATION
SERI
ENZYME HYDROLYSIS
800 SECTION
ANAEROBIC DIGESTION
14667
FIG. 5.1-9
2



R-904 LIGNIN CENTRIFUGE
Q-901 LIGNIN DAY BIN (13'-6" x 52')
G-901 MULTICLONE
T-901 PRIMARY ECONOMIZER
T-903 AIR PREHEATER
G-902 BAGHOUSE
R-902 INDUCED DRAFT FAN
G-903 STACK
W-903 LIGNIN DAY BIN SCREW FEEDER
W-901 BOILER SCREW FEEDER
B-901 LIGNIN BOILER
T-902 SUPERHEATER
R-901 PRIMARY AIR FAN
W-902 ASH SILO BELT CONVEYOR
Q-902 ASH SILO (5' x 18')
W-904 SUPPLEMENTAL WOOD CHIP BELT CONVEYOR
W-905 BOILER WOOD CHIP SCREW FEEDER
TO WASTE TREATMENT (1200 SECTION)
BFW
H2SO4
LIGNIN FROM WATER/ALKALI WASH (200 SECTION)
LIGNIN FROM HYDROLYSIS WATER WASH (400 SECTION)
BOILER BLOWDOWN TO WASTE TREATMENT
H.P. STEAM
METHANE-RICH GAS FROM ANAEROBIC DIGESTION (800 SECTION)
SUPPLEMENTAL WOOD CHIPS FROM WOOD FEEDING SYSTEM (1000 SECTION)
FUEL OIL FROM TRUCK
AIR
P-901 FUEL OIL UNLOADING PUMP
Q-904 NO.2 FUEL OIL STORAGE TANK
P-902 FUEL OIL PUMP
LEGEND: TEMPERATURE, °F
PRESSURE, PSIA
STREAM NUMBER
A-C&D INDICATES 3 OPERATING, 1 SPARE
FINAL
FINAL DRAFT
STONE & WEBSTER ENGINEERING CORPORATION
SERI
ENZYME HYDROLYSIS
900 SECTION
BOILER
14667
FIG. 5.1-10
2

G-1001
TRUCK SCALE

G-1003
TRUCK TRAILER DUMPER

Q-1001
TRUCK RECEIVING HOPPER

G-1004
TRUCK RECEIVING YARD PIT

W-1001
UNLOADING BIN DRAG CHAIN

W-1002
SCALPING SCREEN FEED DRAG CHAIN

G-1002
FRONT-END LOADER

G-1006
PRIMARY MAGNETIC SEPARATOR

G-1005
SCALPING SCREEN

W-1003
DOUBLE WING BELT STACKER

W-1007A,B
RECLAIMER BELT CONVEYOR

W-1004
STACKER FEED CONVEYOR

G-1008
STONE TRAP

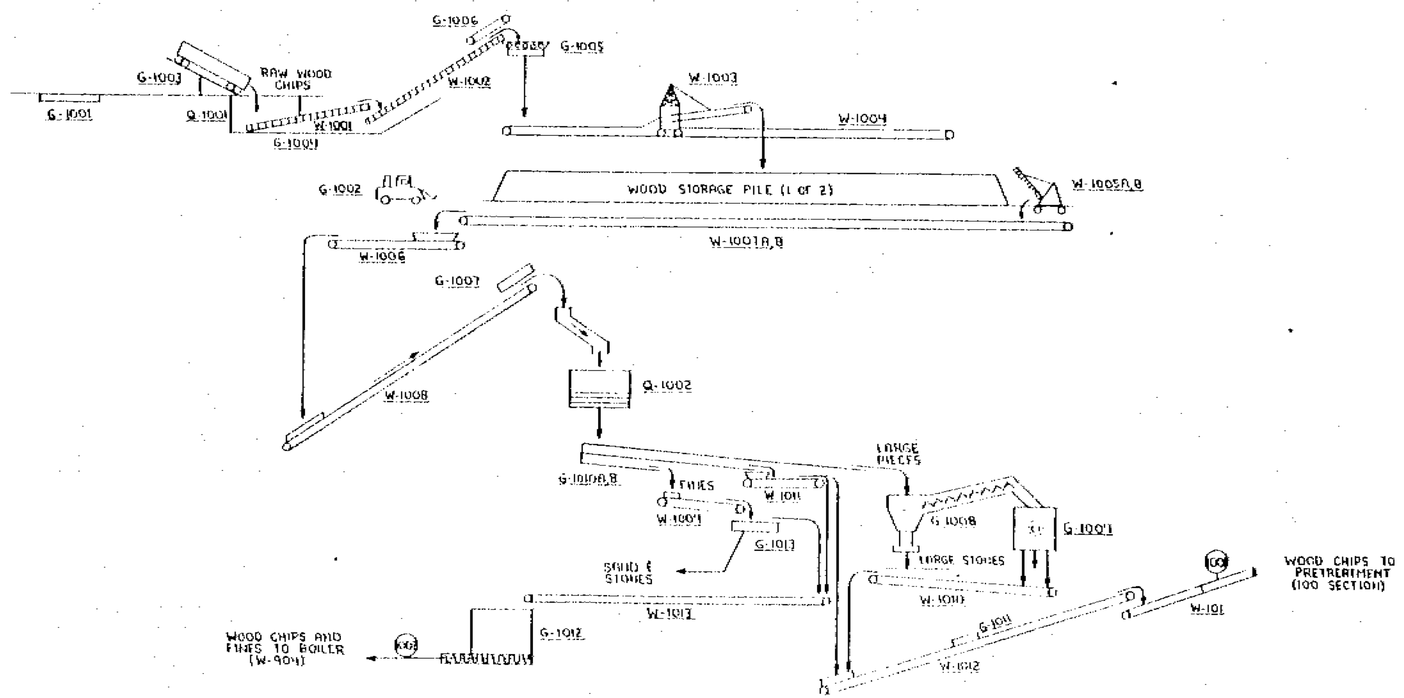
W-1010
LARGE CHIP TRANSFER CONVEYOR

W-1005A,B
TRAVELING SCRAPER/ RECLAIMER

G-1009
OVERSIZED WOOD CHIPPER

G-1011
BELT CONVEYOR SCALE

W-1012
CLEAN WOOD CHIP TRANSFER CONVEYOR



LEGEND: ⬭ TEMPERATURE, °F
⬭ PRESSURE, PSIA
O STREAM NUMBER
A-C&D INDICATES 3 OPERATING, 1 SPARE

W-1013
BOILER FUEL TRANSFER CONVEYOR

G-1012
CHIP STORAGE SILO

W-1006
STORAGE PILE TRANSFER CONVEYOR

W-1008
WOOD CHIP ELEVATING CONVEYOR

G-1007
SECONDARY MAGNETIC SEPARATOR

G-1010A,B
THREE DECK CHIP SCREENS (M, 400 FT²)

Q-1002
SURGE BIN (800 FT³)

W-1009
FINES TRANSFER CONVEYOR

W-1011
MEDIUM CHIP TRANSFER CONVEYOR

G-1013
VIBRATING SCREEN

2		FINAL	LFS
1		FINAL DRAFT	LFS

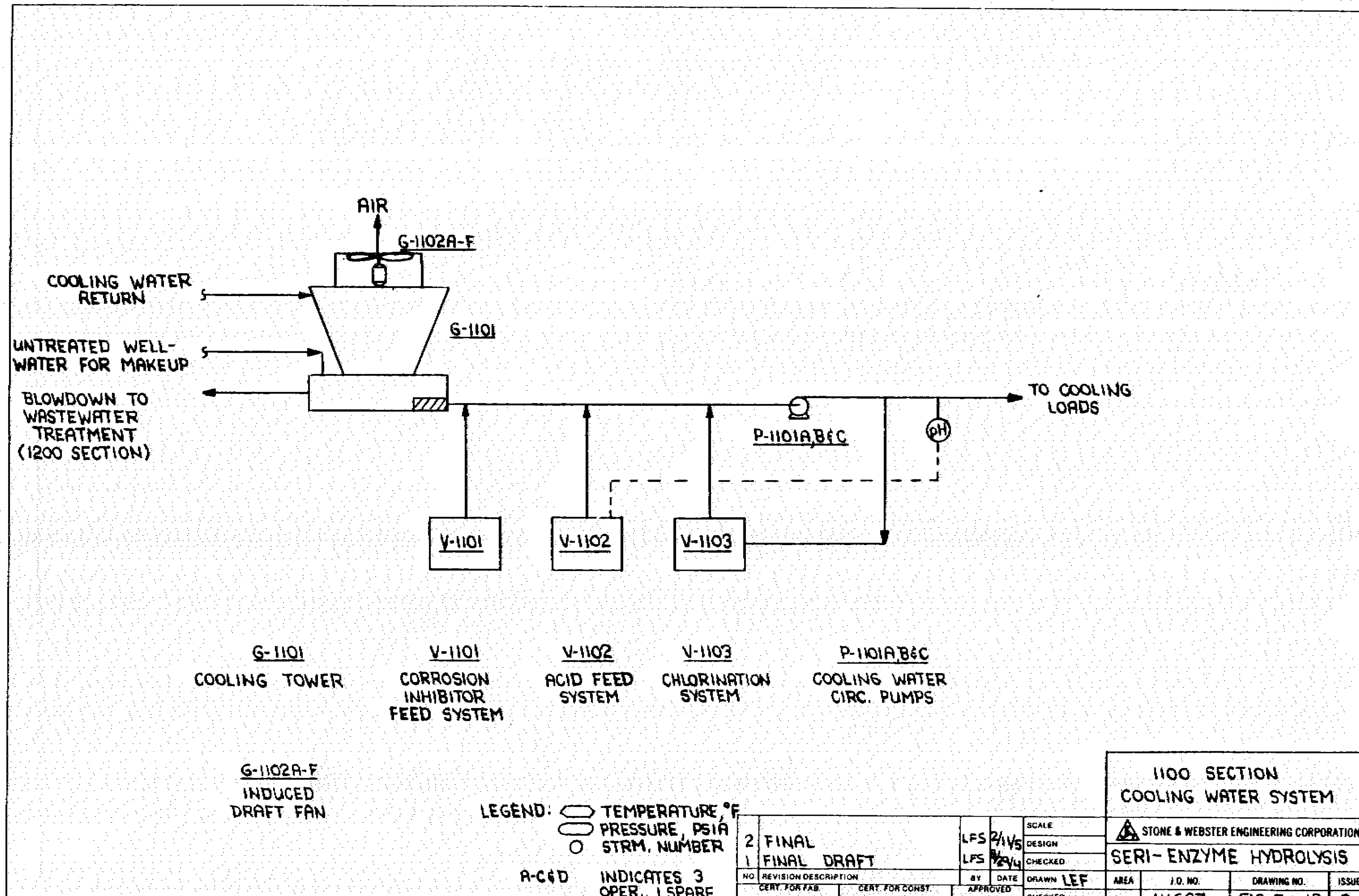
STONE & WEBSTER ENGINEERING CORPORATION

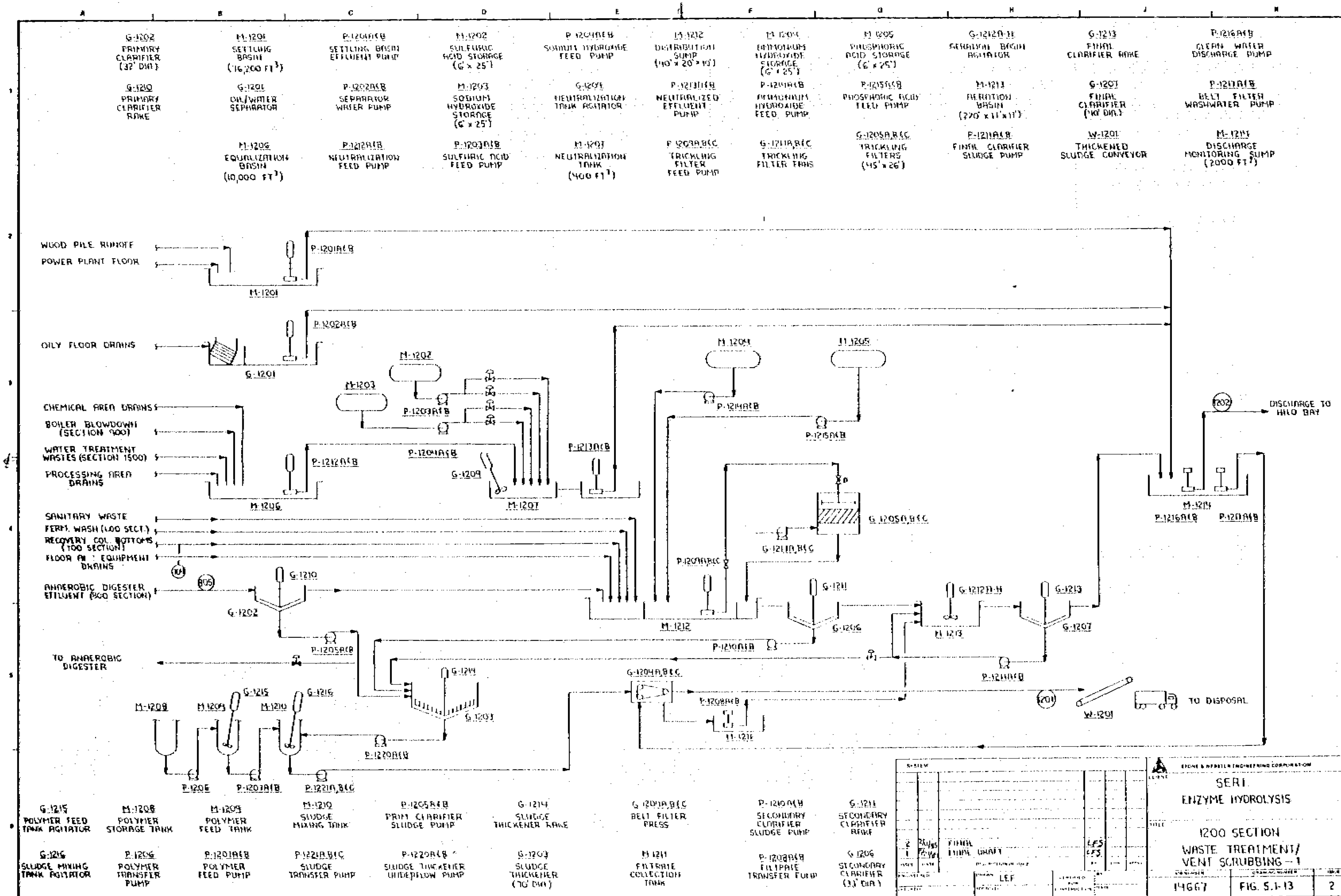
SERI
ENZYME HYDROLYSIS

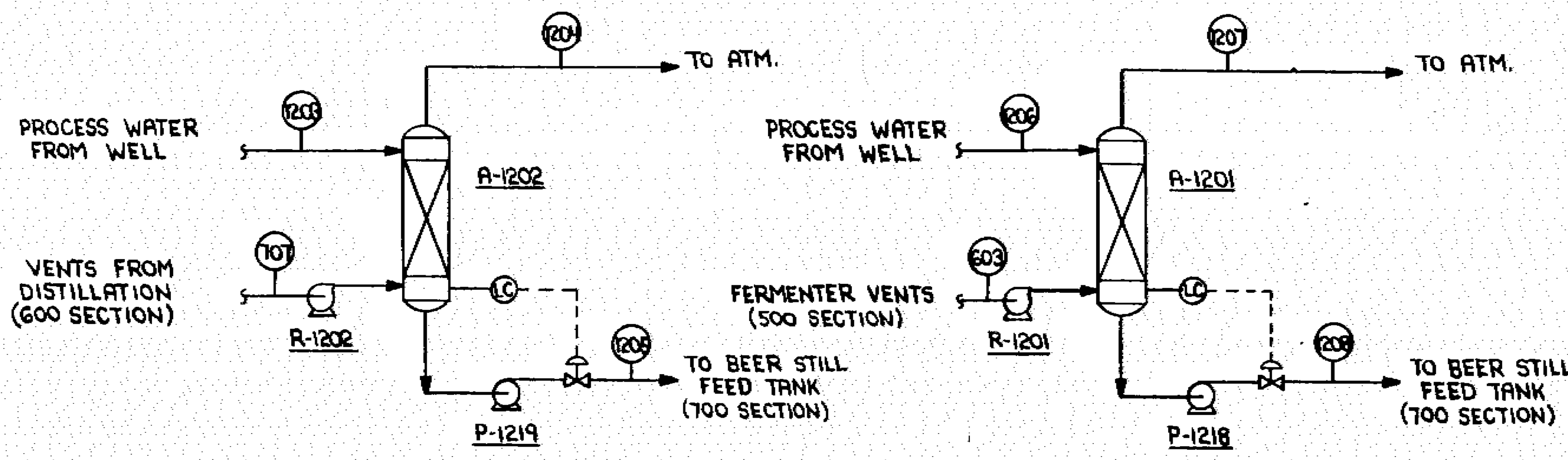
1000 SECTION
FEEDSTOCK HANDLING

LEF

14667 | FIG 5.1-11 | 2


AIR
G-1102A-F
COOLING WATER RETURN
G-1101
UNTREATED WELL-WATER FOR MAKEUP
BLOWDOWN TO WASTEWATER TREATMENT (1200 SECTION)
TO COOLING LOADS
pH
P-1101A,B&C
V-1101
V-1102
V-1103
G-1101
COOLING TOWER
V-1101
CORROSION INHIBITOR FEED SYSTEM
V-1102
ACID FEED SYSTEM
V-1103
CHLORINATION SYSTEM
P-1101A,B&C
COOLING WATER CIRC. PUMPS
G-1102A-F
INDUCED DRAFT FAN
LEGEND:
TEMPERATURE, °F
PRESSURE, PSIA
STRM. NUMBER
A-C&D INDICATES 3 OPER., 1 SPARE
2 FINAL LFS 2/1/85
1 FINAL DRAFT LFS 8/29/84
NO. REVISION DESCRIPTION BY DATE
CERT. FOR FAB. CERT. FOR CONST. APPROVED
SCALE
DESIGN
CHECKED
DRAWN LEF
CHECKED
1100 SECTION
COOLING WATER SYSTEM
STONE & WEBSTER ENGINEERING CORPORATION
SERI- ENZYME HYDROLYSIS
AREA
J.O. NO. 14667
DRAWING NO. FIG. 5.1-12
ISSUE 2


G-1202 PRIMARY CLARIFIER (32' DIA)
G-1210 PRIMARY CLARIFIER RAKE
M-1201 SETTLING BASIN (16,200 FT³)
G-1201 OIL/WATER SEPARATOR
M-1206 EQUALIZATION BASIN (10,000 FT³)
P-1201A&B SETTLING BASIN EFFLUENT PUMP
P-1202A&B SEPARATOR WATER PUMP
P-1212A&B NEUTRALIZATION FEED PUMP
M-1202 SULFURIC ACID STORAGE (6' x 25')
M-1203 SODIUM HYDROXIDE STORAGE (6' x 25')
P-1203A&B SULFURIC ACID FEED PUMP
P-1204A&B SODIUM HYDROXIDE FEED PUMP
G-1209 NEUTRALIZATION TANK AGITATOR
M-1207 NEUTRALIZATION TANK (400 FT³)
M-1212 DISTRIBUTION SUMP (40' x 20' x 10')
P-1213A&B NEUTRALIZED EFFLUENT PUMP
P-1209A,B&C TRICKLING FILTER FEED PUMP
M-1204 AMMONIUM HYDROXIDE STORAGE (6' x 25')
P-1214A&B AMMONIUM HYDROXIDE FEED PUMP
G-1211A,B&C TRICKLING FILTER FANS
M-1205 PHOSPHORIC ACID STORAGE (6' x 25')
P-1215A&B PHOSPHORIC ACID FEED PUMP
G-1205A,B&C TRICKLING FILTERS (45' x 26')
G-1212A-H AERATION BASIN AGITATOR
M-1213 AERATION BASIN (220' x 11' x 11')
P-1211A&B FINAL CLARIFIER SLUDGE PUMP
G-1213 FINAL CLARIFIER RAKE
G-1207 FINAL CLARIFIER (40' DIA)
W-1201 THICKENED SLUDGE CONVEYOR
P-1216A&B CLEAN WATER DISCHARGE PUMP
P-1217A&B BELT FILTER WASHWATER PUMP
M-1214 DISCHARGE MONITORING SUMP (2000 FT³)
WOOD PILE RUNOFF
POWER PLANT FLOOR
OILY FLOOR DRAINS
CHEMICAL AREA DRAINS
BOILER BLOWDOWN (SECTION 900)
WATER TREATMENT WASTES (SECTION 1500)
PROCESSING AREA DRAINS
SANITARY WASTE
FERM. WASH (400 SECT.)
RECOVERY COL. BOTTOMS (700 SECTION)
FLOOR & EQUIPMENT DRAINS
ANAEROBIC DIGESTER EFFLUENT (800 SECTION)
TO ANAEROBIC DIGESTER
DISCHARGE TO HILO BAY
TO DISPOSAL
P-1201A&B
P-1202A&B
P-1212A&B
P-1203A&B
P-1204A&B
P-1213A&B
P-1214A&B
P-1215A&B
P-1209A,B&C
G-1205A,B&C
G-1211A,B&C
G-1212A-H
P-1210A&B
P-1211A&B
P-1216A&B
P-1217A&B
P-1205A&B
P-1208A&B
G-1204A,B&C
P-1220A&B
P-1206
P-1207A&B
P-1221A,B&C
G-1215
G-1216
G-1214
G-1203
M-1208
M-1209
M-1210
M-1211
1202
1201
G-1215 POLYMER FEED TANK AGITATOR
G-1216 SLUDGE MIXING TANK AGITATOR
M-1208 POLYMER STORAGE TANK
P-1206 POLYMER TRANSFER PUMP
M-1209 POLYMER FEED TANK
P-1207A&B POLYMER FEED PUMP
M-1210 SLUDGE MIXING TANK
P-1221A,B&C SLUDGE TRANSFER PUMP
P-1205A&B PRIM CLARIFIER SLUDGE PUMP
P-1220A&B SLUDGE THICKENER UNDERFLOW PUMP
G-1214 SLUDGE THICKENER RAKE
G-1203 SLUDGE THICKENER (70' DIA)
G-1204A,B&C BELT FILTER PRESS
M-1211 FILTRATE COLLECTION TANK
P-1210A&B SECONDARY CLARIFIER SLUDGE PUMP
P-1208A&B FILTRATE TRANSFER PUMP
G-1211 SECONDARY CLARIFIER RAKE
G-1206 SECONDARY CLARIFIER (33' DIA)
2 FINAL
1 FINAL DRAFT
LFS
LFS
LEF
STONE & WEBSTER ENGINEERING CORPORATION
SERI
ENZYME HYDROLYSIS
1200 SECTION
WASTE TREATMENT/
VENT SCRUBBING – 1
14667
FIG. 5.1-13
2



R-1202	A-1202	P-1219	R-1201	A-1201	P-1218
VENT SCRUBBER BLOWER	VENT SYSTEM SCRUBBER (1' x 17'-6")	VENT SCRUBBER PUMP	CO_2 WASH COLUMN BLOWER	CO_2 WASH COLUMN (3'-6" x 16'-6")	CO_2 WASH COLUMN PUMP

LEGEND: ⬭ TEMPERATURE, °F
⬭ PRESSURE, PSIA
○ STREAM NUMBER

A-C&D INDICATES 3 OPERATING, 1 SPARE

NO	REVISION DESCRIPTION	BY	DATE
2	FINAL	LFS	2/11/5
1	FINAL DRAFT	LFS	8/29/84

CERT. FOR FAB. | CERT. FOR CONST. | APPROVED

SCALE | DESIGN | CHECKED | DRAWN LEF | CHECKED

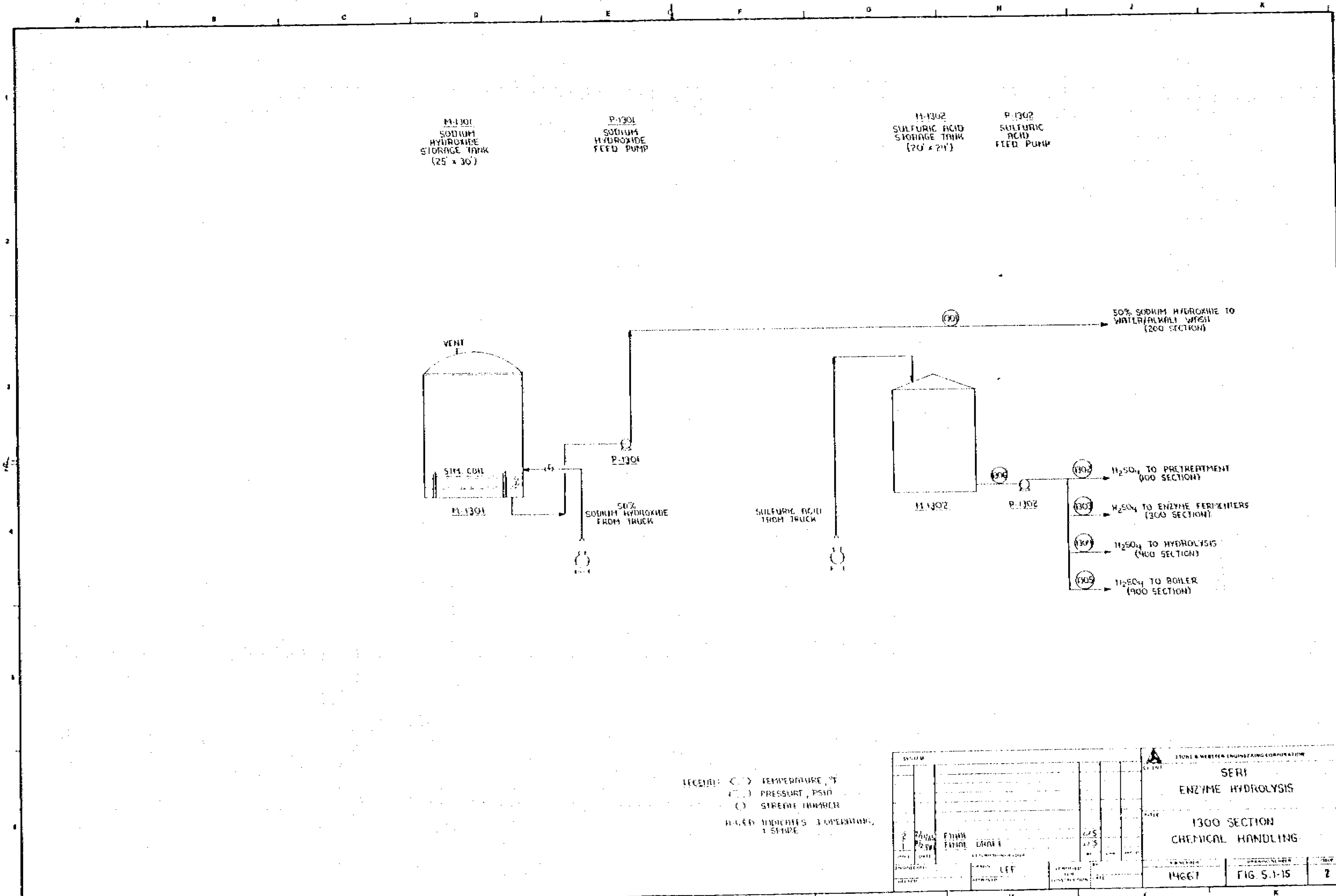
1200 SECTION

WASTE TREATMENT/ VENT SCRUBBING-2

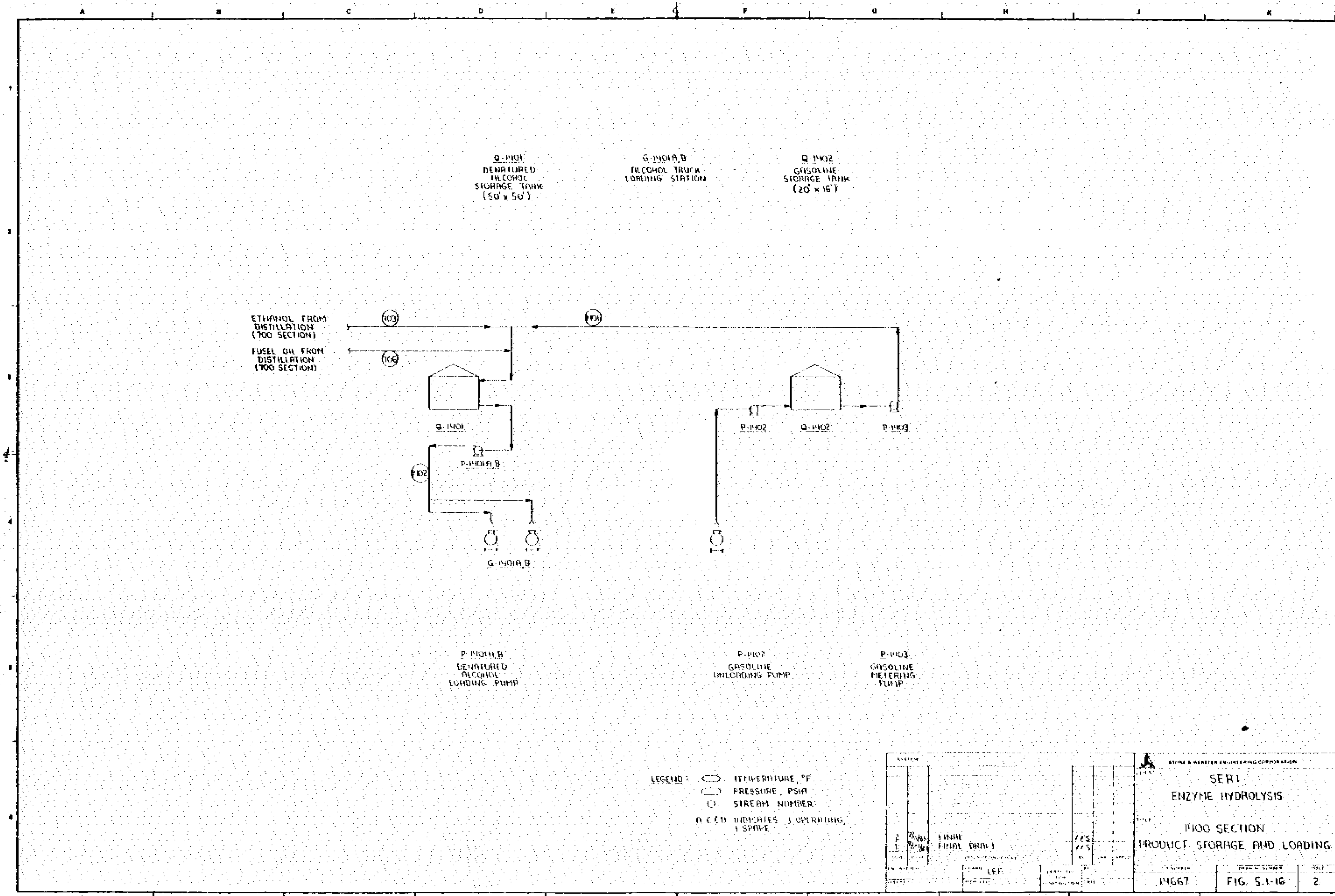
STONE & WEBSTER ENGINEERING CORPORATION

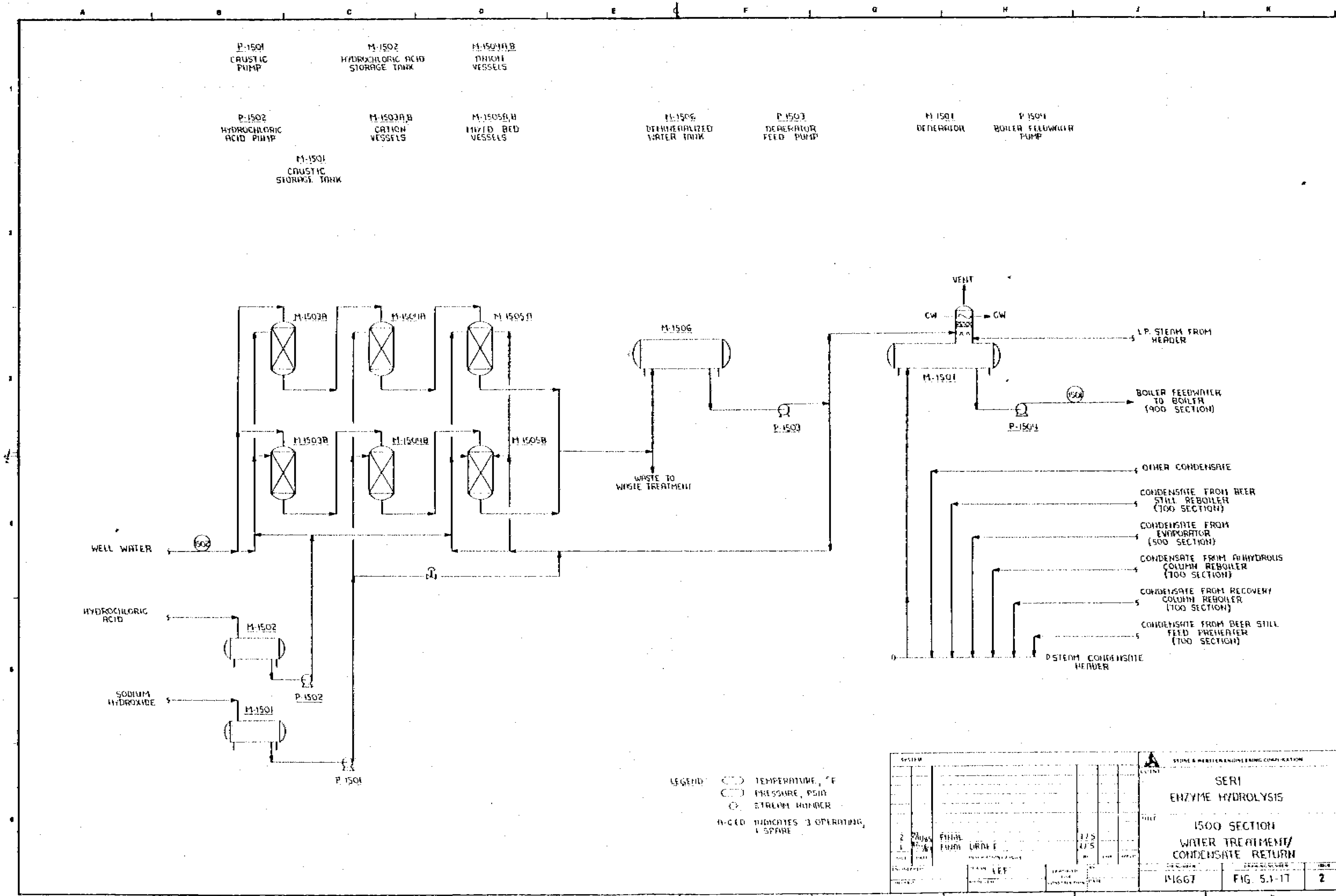
SERI-ENZYME HYDROLYSIS

AREA	J.O. NO.	DRAWING NO.	ISSUE
	14667	FIG. 5.1-14	2



M-1301
SODIUM HYDROXIDE STORAGE TANK
(25' x 30')
P-1301
SODIUM HYDROXIDE FEED PUMP
M-1302
SULFURIC ACID STORAGE TANK
(20' x 24')
P-1302
SULFURIC ACID FEED PUMP
1301
50% SODIUM HYDROXIDE TO WATER/ALKALI WASH
(200 SECTION)
VENT
STM. COIL
M-1301
P-1301
50% SODIUM HYDROXIDE FROM TRUCK
SULFURIC ACID FROM TRUCK
M-1302
1306
P-1302
1302
H_2SO_4 TO PRETREATMENT
(100 SECTION)
1303
H_2SO_4 TO ENZYME FERMENTERS
(300 SECTION)
1304
H_2SO_4 TO HYDROLYSIS
(400 SECTION)
1305
H_2SO_4 TO BOILER
(900 SECTION)
LEGEND: ( ) TEMPERATURE, °F
( ) PRESSURE, PSIA
( ) STREAM NUMBER
STONE & WEBSTER ENGINEERING CORPORATION
SERI
ENZYME HYDROLYSIS
1300 SECTION
CHEMICAL HANDLING
FINAL
FINAL DRAFT
LFF
14667
FIG. 5.1-15
2


Q-1901
DENATURED
ALCOHOL
STORAGE TANK
(50' x 50')
G-1901A,B
ALCOHOL TRUCK
LOADING STATION
Q-1902
GASOLINE
STORAGE TANK
(20' x 16')
ETHANOL FROM
DISTILLATION
(700 SECTION)
FUSEL OIL FROM
DISTILLATION
(700 SECTION)
103
104
106
102
Q-1901
P-1901A,B
G-1901A,B
P-1902
Q-1902
P-1903
P-1901A,B
DENATURED
ALCOHOL
LOADING PUMP
P-1902
GASOLINE
UNLOADING PUMP
P-1903
GASOLINE
METERING
PUMP
LEGEND:
TEMPERATURE, °F
PRESSURE, PSIA
STREAM NUMBER
A,C,E,D INDICATES 1 OPERATING, 1 SPARE
2 FINAL
1 FINAL DRAFT
LEF
STONE & WEBSTER ENGINEERING CORPORATION
SERI
ENZYME HYDROLYSIS
1900 SECTION
PRODUCT STORAGE AND LOADING
14667
FIG. 5.1-16
2



P-1501
CAUSTIC PUMP
M-1502
HYDROCHLORIC ACID STORAGE TANK
M-1504A,B
ANION VESSELS
P-1502
HYDROCHLORIC ACID PUMP
M-1503A,B
CATION VESSELS
M-1505A,B
MIXED BED VESSELS
M-1506
DEMINERALIZED WATER TANK
P-1503
DEAERATOR FEED PUMP
M-1501
DEAERATOR
P-1504
BOILER FEEDWATER PUMP
M-1501
CAUSTIC STORAGE TANK
VENT
CW
CW
L.P. STEAM FROM HEADER
M-1503A
M-1504A
M-1505A
M-1506
M-1501
BOILER FEEDWATER TO BOILER (900 SECTION)
P-1503
P-1504
M-1503B
M-1504B
M-1505B
WASTE TO WASTE TREATMENT
OTHER CONDENSATE
CONDENSATE FROM BEER STILL REBOILER (700 SECTION)
CONDENSATE FROM EVAPORATOR (500 SECTION)
CONDENSATE FROM ANHYDROUS COLUMN REBOILER (700 SECTION)
CONDENSATE FROM RECOVERY COLUMN REBOILER (700 SECTION)
CONDENSATE FROM BEER STILL FEED PREHEATER (700 SECTION)
STEAM CONDENSATE HEADER
WELL WATER
HYDROCHLORIC ACID
M-1502
P-1502
SODIUM HYDROXIDE
M-1501
P-1501
LEGEND:
TEMPERATURE, °F
PRESSURE, PSIA
STREAM NUMBER
A-C-E-D INDICATES 3 OPERATING, 1 SPARE
SYSTEM
2
1
FINAL
FINAL DRAFT
LEF
STONE & WEBSTER ENGINEERING CORPORATION
SERI
ENZYME HYDROLYSIS
1500 SECTION
WATER TREATMENT/
CONDENSATE RETURN
14667
FIG. 5.1-17
2

5.3 OFFSITE FACILITIES

Plant roads, parking areas, and perimeter fencing around the site are shown on the Site Plan (Figure 5.3-1). These items are reflected in the cost estimate. In addition, an allowance was made in the cost estimate for landscaping.

Buildings included in the developed capital costs for this facility are listed in Table 5.3-1.

The costs for furnishings and laboratory equipment are included as an allowance.

TABLE 5.3-1

BUILDING SUMMARY

Building Number	Description	Floor Area (ft^2)	Height (ft)	Comments
1	Administration	6,600	12	Air Conditioned
2	Maintenance/Warehouse	12,800	24/16	Shop has 24' ceiling
3	Service	6,400	9	Air Conditioned
4	Gate House	180	8	Air Conditioned
100	Feed Preparation	5,500	60	Open Sides
200	Steam Explosion Feed Bins Canopy	1,600		Supported by bin support structure
600	Process Air Compressor Shed	1,200	16	
800	Methane Compressor Shed	2,400	16	
1100	Cooling Tower Chlorine Storage	260	10	
1200	Dewatering Equipment Building	5,600	25	
1400	Truck Loading Metal Deck Roof	400	18	Open Sides
1600	Fire Pump and Foam House	300	10	
1601	Fire Water Pump House	300	10	
1602	Instrument/Service Air Compressor Shed	1,200	16	

5.2 UTILITIES

5.2.1 Steam System

The steam system is designed to combust the cellulose and lignin wastes, methane from the anaerobic digester, and supplemental wood to produce high pressure (650 psig) steam. Excess HP steam is let down through back pressure turbines which drive the air compressors (R-301A,B) and provide the required LP steam for the process.

Figure 5.1-10 is a process flow diagram of the steam system. The waste lignin stream from the lignin centrifuge (R-904) is conveyed at 38-percent solids into the boiler. The solids are then fed into the boiler by a screw conveyor (W-901). Methane is fed directly into the boiler bed via a sparger distribution system. Supplemental wood chips are conveyed from the feedstock handling system (Section 1000) and screw fed into the boiler via the boiler wood chip screw feeder (W-905). Fuel oil service is supplied in the boiler section for startup purposes.

The boiler is specified as a fluidized bed type to allow the combustion of the lignin and other wastes without additional drying. The boiler island includes the following fan-to-fan equipment:

<u>Boiler Island Equipment</u>

B-901	Fluidized Bed Boiler
W-901	Boiler Screw Feeder
T-901, T-902, T-903	Boiler Heat Exchange Surfaces
R-901, R-902	FD and ID fans
G-901	Multiclones
G-902	Baghouse

Air for fluidization and combustion is supplied by the primary air fan (R-901). The air is preheated with flue gas in the air preheater (T-903) and enters the boiler (B-901) in the distribution chamber. The fluidized bed of inert material creates a turbulent mixing to thoroughly combust the lignin and waste particles. Particles carried overhead with the flue gas enhance the heat transfer coefficients in the heat exchanger banks.

The larger particles are captured in the multiclone (G-901) and are recycled to the bed. The fine particles, primarily wood ash, are sent to the baghouse (G-902), where they are removed from the flue gas. Wood ash exits the baghouse and is transported by the belt conveyor (W-902) to the ash storage silo (Q-902), which is sized for one-day storage. Waste ash is taken from the site by truck.

The steam header diagram is shown in Figure 5.2-1. The major uses include the steam explosion guns (V-203A-D), the evaporator, and the distillation reboilers (T-719, T-708, T-709).

High pressure steam supplied by the boiler (B-901) is desuperheated and used in the steam explosion guns (V-203A-D). The air compressors (R-301A,B) are driven by HP steam through back pressure turbines providing LP steam at 50 psig. The clean condensate is returned to the deaerator (M-1507).

5.2.2 Cooling Water System

The cooling water system is shown in Figure 5.1-12. The cooling water summary for the enzyme-based ethanol plant is given in Figure 5.2-2. The function of this system is to supply cooling water to the individual component coolers and heat exchangers located throughout the plant and to reject this waste heat to the atmosphere by both sensible and evaporative heat transfer in the cooling tower. A closed-cycle cooling system was specified which utilizes a wet, mechanical draft cooling tower to dissipate heat to the atmosphere.

The study design basis uses a summer wet bulb temperature of 75°F and a maximum cold water temperature of 85°F, thus dictating a design cooling tower approach of 10°F. The cooling tower design drift rate is specified not to exceed 0.001 percent of the cooling water system flow.

The cooling water circulating pumps each have a capacity of 6,500 gpm at 100 ft total discharge head (TDH). Three 50-percent pumps are provided. The distribution piping system is sized so that, at design flow, the velocity is approximately 10 fps. Two of the three 50-percent capacity cooling water pumps (P-110A,B&C) normally circulate the cooling water through carbon steel distribution pipes. Hot water is returned to the distribution trough of the multi-cell cooling tower (G-1101). Water spills downward through the tower packing while airflow is induced upward by the tower fans. The falling water is cooled by both evaporative and sensible heat transfer to the air and is collected in the basin for recirculation.

Biofouling in the cooling water system and related exchangers is controlled by the periodic injection of chlorine into the cooling water pump inlet. An acid feed system maintains circulating water pH. An inhibitor feed system is provided for controlling corrosion. Untreated well water is used for cooling tower makeup (275 gpm).

The refrigerated water system supplies chilled water to areas of the plant which must be cooled below a reasonable approach for the cooling water system. The package refrigeration unit is shown schematically in Figure 5.1-6 (T-602, T-603, R-602).

The refrigerated water summary is given in Figure 5.2-2. The primary uses are the fermenter feed chiller (T-604), the fermenter refrigeration loop cooler (T-602), and the distillation vent condensers (T-726, T-

727). A cooling water inlet temperature of 60°F was specified with a 10°F rise giving a total refrigerated water flow rate of 1197 gpm.

5.2.3 Instrument/Service Air System

The Instrument/Service Air System is designed to provide compressed air to the process plant and offsites. Raw, compressed air is supplied to the service air header, while clean, dry air is required for the instrument and control header. Two compressors (R1601, R1602) are specified to supply 600 scfm each at 100 psig. In normal operation the compressors are isolated, each connected to one of the two main air headers. Special valving is provided to allow one compressor to back-up the other in the event of equipment failure or required maintenance.

Both compressors are equipped with filter/silencers which remove 98 percent of the particles greater than 3 microns. In addition, the instrument air system contains pre- and after-filters (V1603, V1604) around a desiccant dryer system (V1607) to remove essentially 100 percent of particles 3 microns and greater, and to dry the instrument air to a dew point of -40°F.

5.2.4 Fire Protection System

The function of the fire protection system is to detect, annunciate, suppress, and extinguish any fire on the plant site or in any building or enclosed area, either automatically and/or manually.

The fire protection system is designed, installed, and tested in accordance with the following National Fire Protection Association (NFPA) Standards:

NFPA-11	Foam Extinguishing Systems
NFPA-12A	Halon 1301 Systems
NFPA-13	Sprinkler Systems Installation
NFPA-14	Standpipe and Hose Systems
NFPA-20	Centrifugal Fire Pumps
NFPA-24	Outside Protection
NFPA-70	National Electrical Code
NFPA-214	Water Cooling Towers

All components shall be listed or approved by Underwriters Laboratories (UL) or Factory Mutual (FM).

The fire protection system is designed to provide 2,500 gpm of water to the plant fire water loop, to provide Halon as required to the plant control room, and to provide foam to storage tanks containing volatile organic compounds.

Site water storage for fire protection is required equaling 2-1/2 hours of operation of the maximum sprinkler demand. One storage tank is provided with a capacity of 400,000 gallons. Two 2,500 gpm fire water pumps are supplied; one is motor-driven (P-1601) and the other is

TABLE 5.2-1

COOLING WATER AND REFRIGERATED WATER EQUIPMENT SUMMARY

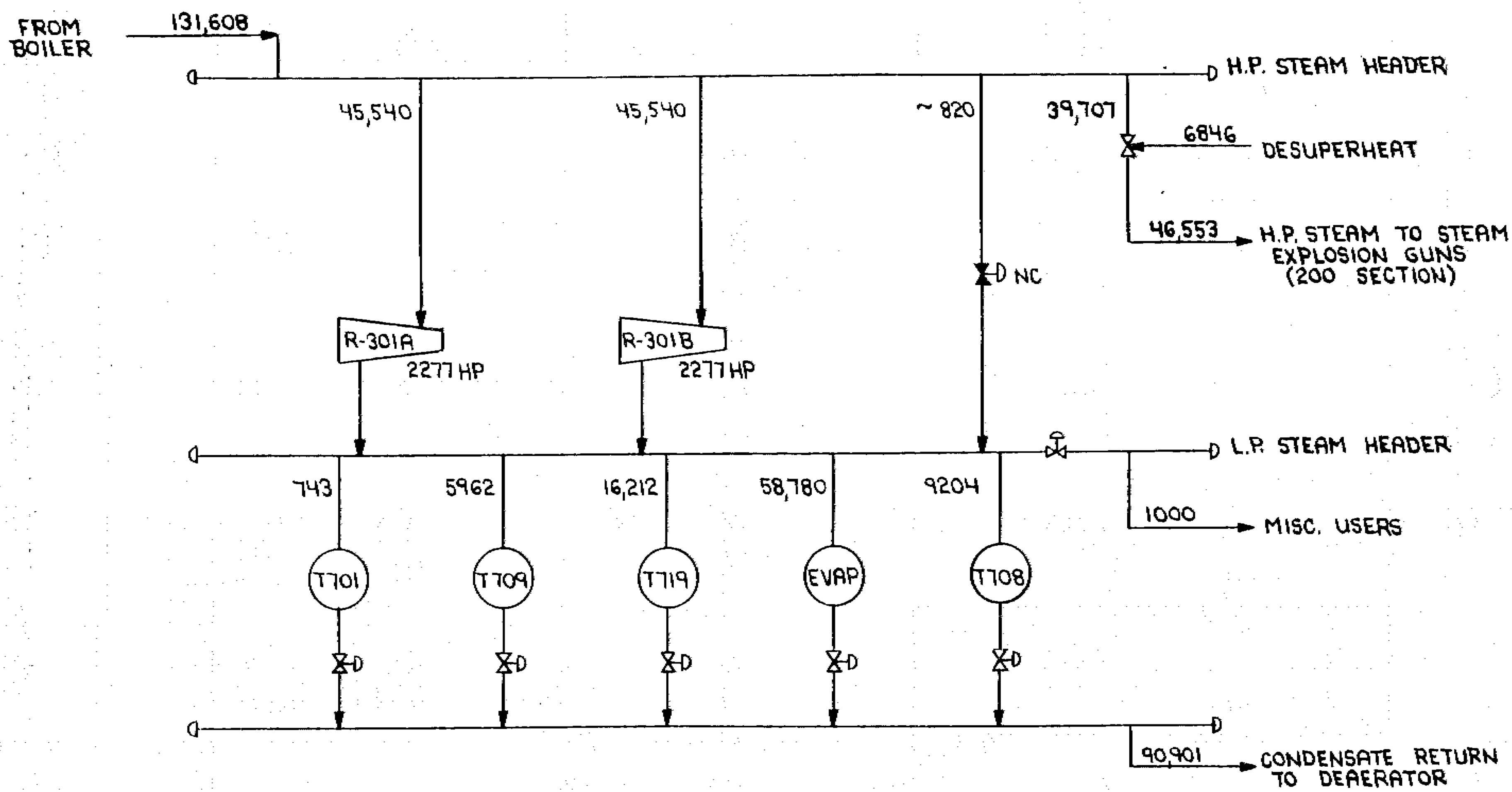
Cooling Water Equipment

Item No.	Service	Total Duty (MM Btu/hr)	T in (°F)	T out (°F)	Total Flow (gpm)
T-201	Vacuum Flash Condenser	18.50	85	110	1,480
T-303A,B	Air Sparge Cooler	5.60	85	110	448
T-304A,B	Air Compressor Intercooler	5.60	85	110	448
T-401	Hydrolysis Dilution Cooler	6.36	85	110	509
T-501	Evaporator Surface Condenser	63.2	85	110	5,056
T-601	Fermenter Feed Cooler	8.00	85	110	640
T-603	Refrigerant Condenser	12.29	85	110	983
T-703	Recovery Column Overhead Condenser	5.27	85	110	422
T-704	Ethanol Product Cooler	1.46	85	105	146
T-705	Beer Still Bottoms Cooler	9.09	85	110	727
T-711	Decanter Feed Cooler	0.96	85	105	96
T-712	Fusel Oil Cooler	0.05	85	110	4
T-713	Anhydrous Column Purge Cooler	0.01	85	110	1
T-715	Anhydrous Column Overhead Condenser	15.40	85	110	1,232
T-717	Anhydrous Column Hold Tank Feed Cooler	0.70	85	110	56
T-724	Anhydrous Column Hold Tank Vent Condenser	0.002	85	110	0.1
T-801	Digester Feed Cooler	3.28	85	110	262

Refrigerated Water Equipment

Item No.	Service	Total Duty (MM Btu/hr)	T in (°F)	T out (°F)	Total Flow (gpm)
T-301A,B	Fermenter No. 1 Recycle Cooler	3.8	60	70	760
T-302A,B	Fermenter No. 2 Recycle Cooler	2.44	60	70	488
T-602	Refrigeration Loop Cooler	4.13	60	70	826
T-604	Fermenter Feed Chiller	3.00	60	70	600
T-726	Beer Still Vent Condenser	0.16	60	70	32
T-727	Degasser Drum Vent Condenser	0.01	60	70	3

STEAM HEADER DIAGRAM

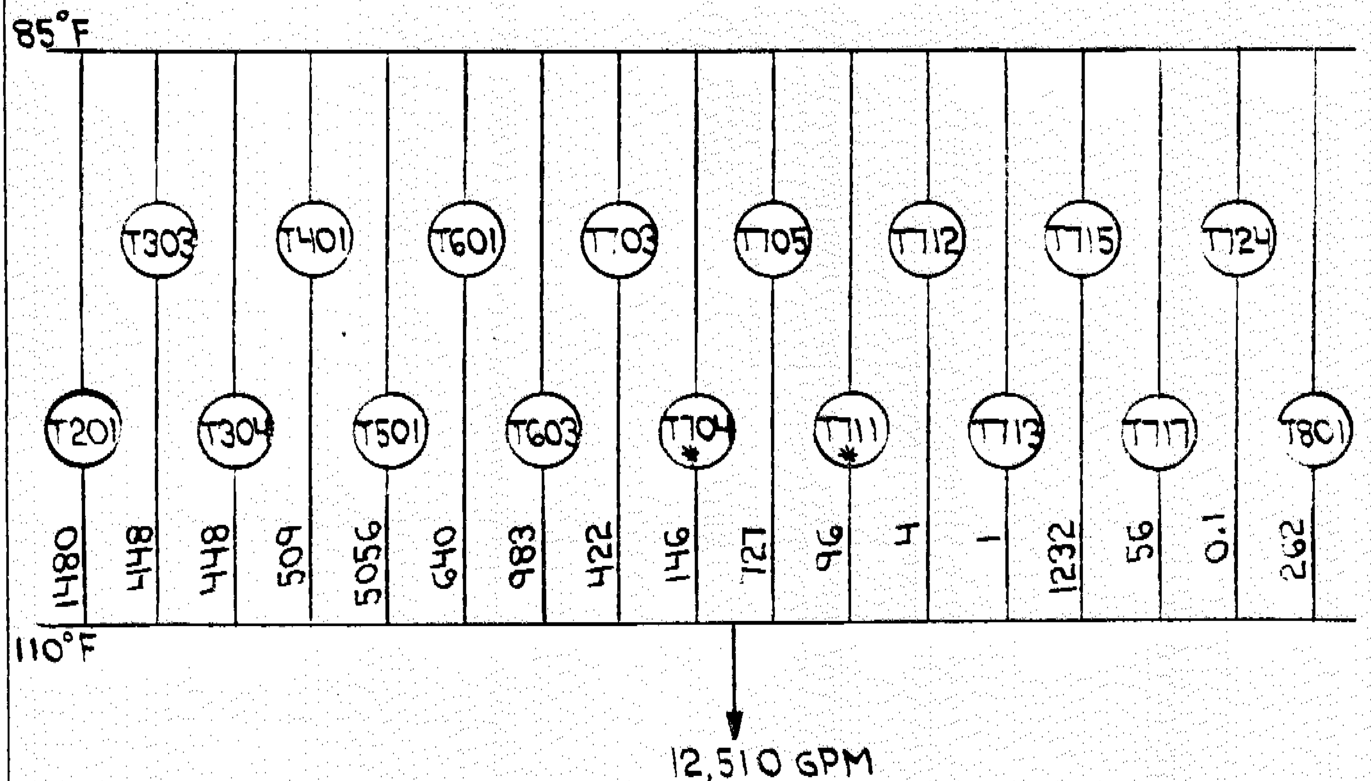


NOTE: FLOWS SHOWN ARE IN LB/HR

NO.	REVISION DESCRIPTION	BY	DATE		
2	FINAL	LFS	2/11/5	SCALE	STONE & WEBSTER ENGINEERING CORPORATION
1	FINAL DRAFT	LFS	8/84	DESIGN	SERI-ENZYME HYDROLYSIS
	CERT. FOR FAB. / CERT. FOR CONST.	APPROVED		CHECKED	
				DRAWN LEF	AREA: — J.O. NO.: 14667 DRAWING NO.: FIGURE 5.2-1 ISSUE: 2
				CHECKED	

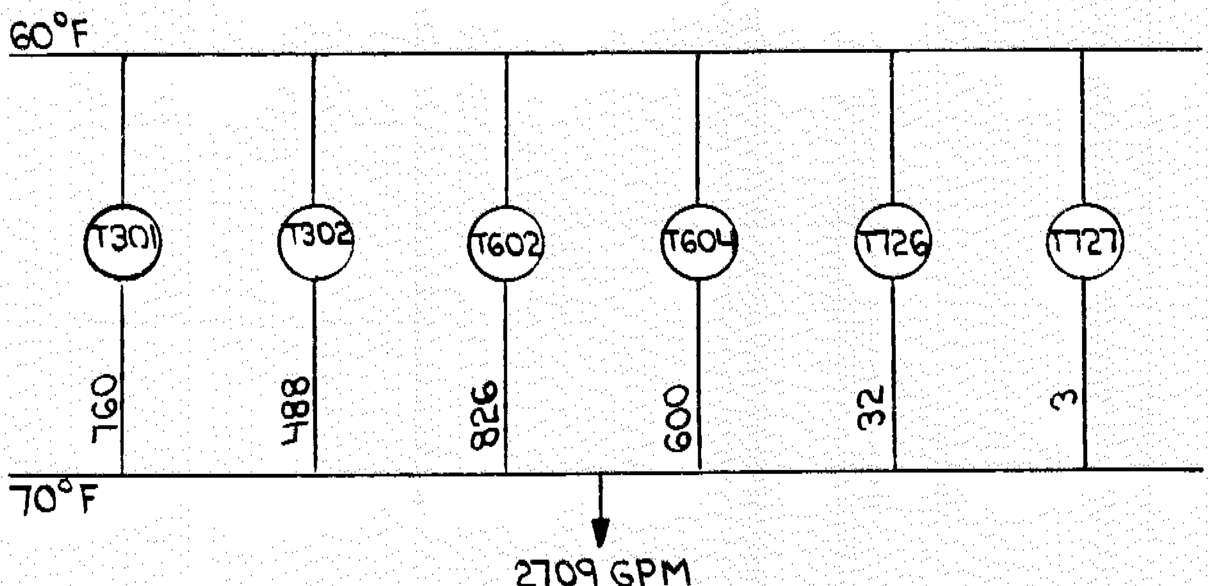
▲ 5038.60B

COOLING WATER SUMMARY



* COOLING WATER OUTLET TEMPERATURE IN T-704, T-711 IS 105°F

REFRIGERATED WATER SUMMARY



NO.	REVISION DESCRIPTION	BY	DATE				
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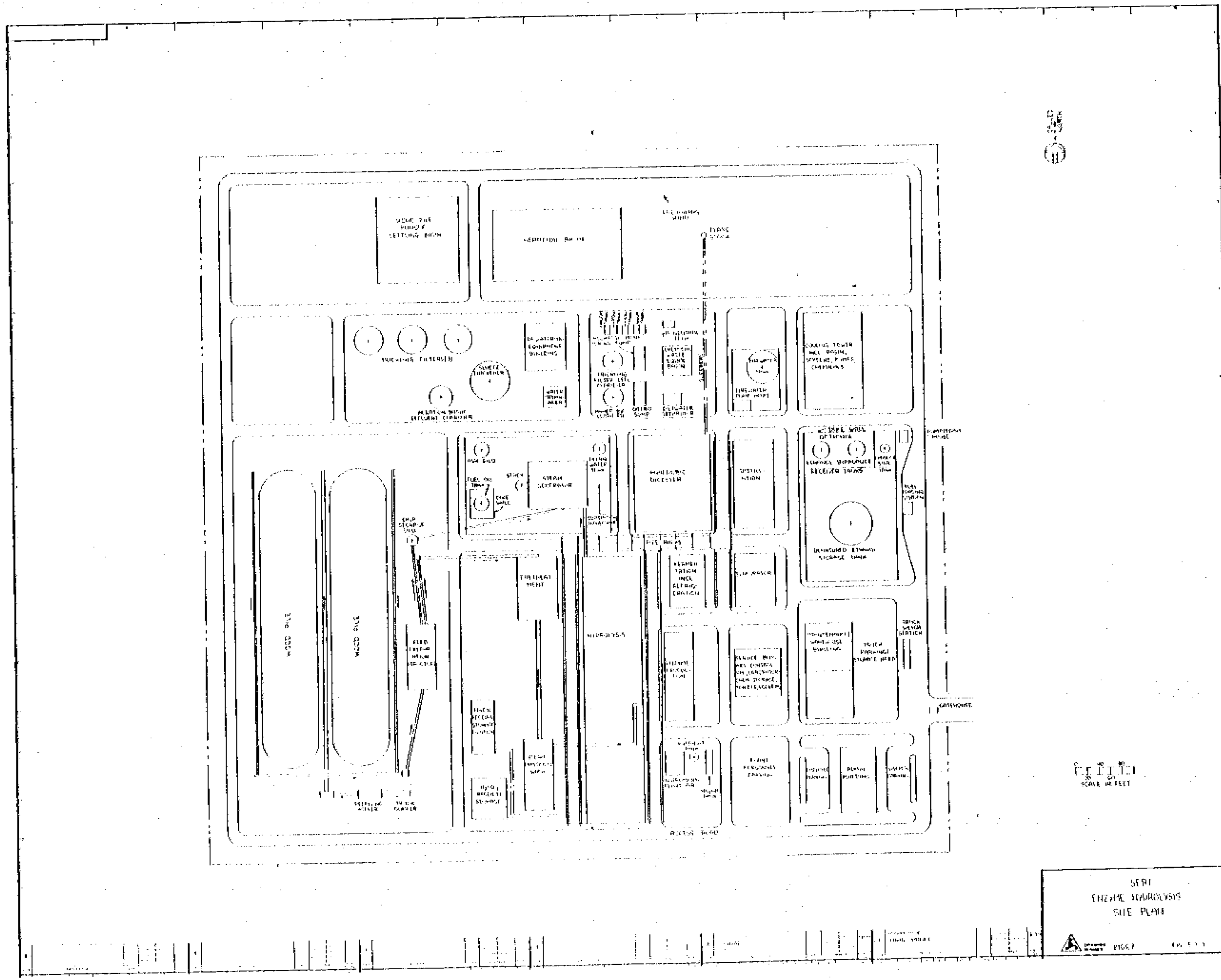
1	FINAL DRAFT	LFS	9/24/84	CHECKED	SERI - ENZYME		
	CERT. FOR FAB. / CERT. FOR CONST.	APPROVED		DRAWN LEF	AREA	J.O. NO. 14667	FIG. 5.2-2
				CHECKED			ISSUE 2


WOOD PILE
WOOD PILE
STEAM GENERATOR
HYDROLYSIS
ACCESS ROAD
SCALE IN FEET
SERI
ENZYME HYDROLYSIS
SITE PLAN

5.4 ENVIRONMENTAL CONSIDERATIONS AND COMPLIANCE

5.4.1 Water Quality Standards and Permit Requirements

Figure 5.4-1 and Table 5.4-1 show the water balance for the entire facility. Based on the annual average requirements, 846 gpm of fresh water are required to support plant operation. Using Haulaula Well as a typical water source, it was assumed that quality and quantity of water available was sufficient to allow use without treatment, except for boiler makeup water requirements. It is recommended that this flow be treated using cation, anion, and mixed bed exchanger to ensure trouble-free boiler and turbine operation. The makeup demineralizer system consists of two 100-percent trains, each sized for approximately 95 gpm.

Cooling water treatment consists of chlorine addition for biofouling control of the main condensers and sulfuric acid feed to control calcium carbonate sealing.

The wastewater treatment system is designed to treat an average flow of 344 gpm from the cogeneration and process facilities. Wastewaters associated with cogeneration plant operation consist of relatively small quantities of floor and equipment drainage, boiler blowdown, and makeup demineralizer regeneration wastes, as well as the larger cooling tower blowdown flow. Effluent limitations for the steam electric generating industry, 40CFR423, may be used as a guide to evaluate requirements applicable to these wastes. At the state level, effluent limitations are generally similar to the required minimum federal limitations. However, additional treatment requirements may be imposed by state authorities to minimize the impact of a discharge on any particular receiving water.

Wastewater associated with the ethanol plant consists of fermenter washwaters, recovery column bottoms, floor and equipment drains, and anaerobic digester effluent, as well as sanitary wastes from the entire facility. No specific effluent limitations have been developed by EPA for the fuel ethanol industry. Since these wastewaters are characterized by conventional pollutants, i.e., biological oxygen demand (BOD) and suspended solids, which are regulated for municipal publically-owned treatment works, the average municipal treatment plant effluent limits defined in 40CFR133 were applied to process wastewaters.

To minimize the size and cost of the various components, waste streams were segregated according to treatment requirements.

Wood pile runoff and floor and equipment drains from the cogeneration area, which contain primarily suspended solids, are collected and held in a settling basin. The basin is sized to hold the flow from the wood pile and materials handling areas resulting from the 100-yr, 24-hr storm, or approximately 570,000 gallons. This design presumes that sufficient detention time can be provided to allow for settling of suspended solids, without additional chemical treatment, to levels permitted for discharge.

Oily drains from the cogeneration area are collected and treated in a corrugated plate-type oil/water separator at a design rate of 50 gpm.

Chemical area drains from both the process and cogeneration areas, boiler blowdown, and makeup demineralizer system regeneration wastes, are collected and the pH adjusted, if necessary, before discharge. Since these flows are intermittent and variable, an equalization train is provided upstream of the neutralization system. The basin is sized to hold a maximum daily flow of approximately 75,000 gallons plus an allowance for freeboard and sludge accumulation. Automatic sulfuric acid and sodium hydroxide feed equipment are provided for neutralization.

The remainder of plant wastes, consisting of sanitary wastes, process area floor and equipment drains, fermenter wash water, recovery column bottoms, and anaerobic digester effluent, are collected and biologically treated before discharge. The average daily flow of the biological treatment system is 287 gpm and contains approximately 14,550 mg/l of COD, 6,500 mg/l of BOD, and 117 mg/l of suspended solids. Due to the high BOD and COD, a two-stage process is required to achieve compliance with the assigned effluent limits. Trickling filters are used as roughing filters to remove approximately 50 percent of the BOD in the influent. Ninty-nine percent of the remaining BOD is removed in an aeration basin.

Before combining the various waste streams in the trickling filter distribution sump, the anaerobic digester effluent is clarified to remove suspended solids. The clarifier is 34 ft in diameter and is designed for a hydraulic load of 0.4 gpm/ft^2.

The first stage of biological treatment consists of three 45 ft diameter, 26 ft deep plastic media packed towers and one 33 ft diameter effluent clarifier. A portion of the trickling filter effluent is recycled back to the filters and the excess portion is discharged to the clarifier for solids removal. The hydraulic loading of 0.7 gpm/ft^2/filter improves the organic removal efficiency by equalizing and diluting the high organic load and provides for efficient distribution of the wastewater on the filter media. Low speed fans are provided for each filter to ensure that aerobic conditions are maintained within the filter. A design organic load of 200 lb BOD/day/1000 ft^3 for 50-percent BOD reduction was used in accordance with previous investigations. Each trickling filter is identical and has a total media volume of approximately 41,000 ft^3.

A phosphoric acid feed system, with pH adjustment using ammonium hydroxide, is utilized to adjust the pH and provide the phosphorous and nitrogen requirements of the trickling filter microorganisms.

The portion of the trickling filter effluent not recycled is treated in a 33-ft-diameter clarifier. A clarifier overflow rate of approximately 0.4 gpm/ft^2 was used. The clarifier effluent is discharged to an aeration basin where the wastewater is biologically stabilized under aerobic conditions. The aeration basin is 220 ft x 117 ft x 17 ft sidewall depth (SWD), based on a BOD loading of 0.2 lb BOD/lb mixed liquor volatile suspended solids (MLVSS) and a resulting aeration period of approximately 7 days. Eight floating low-speed mechanical aerators are provided to supply the dissolved oxygen necessary for process requirements and to maintain microorganisms and other solids in suspension. Wastewater and suspended microorganisms are continuously discharged from the basin and the biological solids are separated from the biologically treated wastewater in a final clarifier. A portion of the settled sludge is returned to the aeration basin to maintain a viable biomass and the remainder is treated for disposal. The 40-ft-diameter aeration basin effluent clarifier is designed at a hydraulic load of 0.3 gpm/ft^2.

Solids removed by the various clarifiers are pumped to a 70 ft diameter sludge thickener, designed for a solids loading of 8 lb/ft^2/day. The thickener increases the solids content of the sludge from 1 percent to 3 percent. The sludge is further dewatered, after sludge conditioning by polymer addition, using three 1.5-meter belt filter process.

The filtrate is returned to the aeration basin and the 20 percent solids sludge is conveyed by truck to an appropriate disposal site. The expected total dry solids sludge production is expected to be approximately 30,000 lb/day.

The total cooling tower blowdown flow rate of 69 gpm is recycled to provide chemical dilution water and wash water makeup for the ethanol plant.

All of the various treated wastewaters are directed to a discharge monitoring sump prior to discharge. State water quality laws require open coastal discharges to be discharged into a minimum of 60 ft of water or 1000 ft offshore, whichever distance is shorter. The combined wastewater will be discharged into Hilo Bay through an offshore discharge structure.

The treatment of process wastewaters results in the production of sludges requiring some form of disposal. Dewatering to levels sufficient for trucking to an appropriate offsite landfill was assumed. Three alternatives are available:

1. Incineration in the fluidized bed boiler

2. Land farming

3. Application to the tree plantation for purposes of fertilization

Land availability, environmental considerations, and location of the various disposal sites must be identified before alternatives No. 2 and 3 are to be considered viable. Alternative No. 1 must consider either sufficient dewatering to permit incineration in the boiler without fuel penalty or minimizing dewatering requirements and assessing fuel penalties. The economics of each alternative would require additional investigation.

A National or State Pollutant Discharge Elimination System (NPDES/SPDES) permit will be required since the facility will discharge treated wastewaters into surface waters. Information submittals will be required by the EPA in order to assess the need for the project consideration of alternatives, and environmental impacts. EPA will then determine if additional information in the form of an Environmental Impact Report or Environmental Information Document is required.

Plans for offsite disposal of solid, nonhazardous wastes should include some form of agreement with operators for the existing land disposal facility licensed by the state.

Other federal and state permits or approvals may be required depending on unique site and project characteristics. Typically, such permits may include, but not be limited to, the following:

- Federal Aviation Regulation (FAR77) relative to stack height clearance for aircraft

- U.S. Army Corps of Engineers, Permits for Structures or Work in Navigable Waterways; Drudge and Fill Permit

- Permits by State Public Utilities Commission and by U.S. Federal Energy Regulation Commission for electrical generation and transmission lines

- State and Local Agency Permits for fuel oil, fuel alcohol storage, and various other building permits

- State approval if site construction is within an approved Coastal Zone Management Plan

5.4.2 Air Quality Standards and Permit Requirements

Hawaii's Department of Health (DOH) has recently proposed to relax its ambient air quality standards for particulates and sulfur dioxide (SO_2) to the same levels as the National Ambient Air Quality Standards (NAAQS) (Table 5.4-2). However, nitrogen dioxide (NO_2), carbon monoxide (CO), and ozone state standards will remain more stringent than the federal standards. Hilo is in compliance with all state and federal ambient air quality standards.

An ethanol plant at Hilo will have to obtain a Prevention of Significant Deterioration (PSD) permit and an Authority to Construct permit from DOH before starting construction. The site is approximately 25 miles from a federal Class I area, the Volcanos National Park, and emissions from the plant must not contribute to ambient air quality concentrations in the Park greater than Class I increments for SO_2 and particulates (Table 5.4-3). Air quality permits require that state and federal emission limits are achieved and that Best Available Technology (BACT) is applied to all sources of significant emissions. The DOH has proposed to define significant emissions equivalent to federal levels (Table 5.4-4). Applicable state emission standards include a particulate limit of 40 lb/hr (0.4 lb/100 lb process weight), an opacity limit of 20 percent, a sulfur in fuel limit of 2 percent, and an ethanol storage requirement for internal floating roof tanks, or similar control equipment. Applicable federal emission limits are comprised of New Source Performance Standards (NSPS) for industrial boilers and for volatile organic compound (VOC) distillation operations and equipment leaks. The industrial boiler and distillation regulations are in the proposal stage but, nevertheless, apply to this project. The NSPS limit boiler particulate emissions to 0.1 lb/10^6 Btu and essentially require process VOC emissions (after all recovery systems) to be reduced by 98 percent, or to a concentration of 20 ppm, whichever is less stringent. An equipment leak detection and repair program for VOC is also required. The requirement for a 98 percent reduction in VOC emissions can be waived if a source maintains a total resource effectiveness (TRE) index of greater than one. The TRE index is a measure of the resources needed for additional reductions in VOC emissions and is based on flow rates, heating values, and corrosive properties of the vent stream.

Emissions and Emission Controls

Particulates from the boiler will be limited to 0.03 lb/10^6 Btu with the application of a baghouse filter (Table 5.4-5). This level of control is more stringent than the state or federal emission limit but is representative of BACT. Nitrogen oxides and CO, which have no specific state or federal limits, will be controlled by proper combustion design, also representative of BACT. The proposed atmospheric fluidized bed boiler yields low NOx emissions due to lower combustion zone temperatures. Emission estimates for the boiler were developed from pilot plant test data and, in the case of particulates, from the baghouse specification (Table 5.4-5). Sulfur dioxide emissions are neglible because the wood-derived fuel contains neglible sulfur. Volatile organic compounds and lead are also emitted in neglible quantities from the boiler.

Process emissions, excluding carbon dioxide and water, consist primarily of gaseous ethanol. The vent stream from the vapor recovery system will be piped to the boiler for combustion to achieve a 90-percent reduction in VOC emissions. The carbon dioxide wash vent stream will be treated with a carbon adsorber to remove VOC to comply with the NSPS for distillation operations.

TABLE 5.4-1

UTILITY WATER BALANCE
ENZYME-BASED ETHANOL PLANT

Stream No.	Description	Average Yearly Flow (gpm)
1	Haulaula Well	846
2	Service Water	25
3	Raw Water	98
4	Demineralized Water	35
5	Potable and Sanitary	3
6	Service Water	25
7	Sterilized Water	7
8	Chemical Dilution	295
9	Process Makeup	155
10	Process Makeup (Total)	224
11	Cooling Tower Makeup	275
12	Wood Pile Runoff	10
13	Floor and Equipment Drains	11
14	Oily Floor Drains	5
15	Chemical Area Drains	10
16	Boiler Blowdown	5
17	Water Treatment Wastes	8
18	Sanitary Waste	3
19	Chemical Area Drains	8
20	Anaerobic Digestor Effluent	260
21	Recovery Column Bottoms	2
22	Process Area Floor Drains	21

TABLE 5.4-1 (Cont)

Stream No.	Description	Average Yearly Flow (gpm)
23	Clean In-Place Wash	1
25	Cooling Tower Blowdown	69
26	Cooling Tower Blowdown (Recycled)	69
27	Evaporation & Drift	206
28	Settling Basin Effluent	21
29	Treated Oily Wastes	5
30	Neutralization System Influent	31
31	Neutralization System Effluent	31
32	Process Wastewater	287
33	Biological Treatment System Effluent	277
34	Biological Sludge	80
35	Filtrate	70
36	Dewatered Sludge (Entrained Water)	10
37	Cooling Tower Blowdown (Disabled)	0
38	Discharge to Hilo Bay	334

TABLE 5.4-2

AMBIENT AIR QUALITY STANDARDS
($\mu g/m^3$)

		Hawaii	NAAQS	
			Primary	Secondary
Sulfur Dioxide	Annual	20	80	-
	24-hour	80	365$^{(1)}$	-
	3-hour	400	-	1,300
Particulates	Annual	55	75$^{(2)}$	60$^{(3)}$
	24-hour	100	260$^{(1)}$	150$^{(1)}$
Nitrogen Dioxide	Annual	70	100	100
	24-hour	150	-	-
Carbon Monoxide	8-hour	5,000	10,000$^{(1)}$	10,000$^{(1)}$
	1-hour	10,000	40,000$^{(1)}$	40,000$^{(1)}$
Lead	Calendar quarter	1.5	1.5	1.5
Ozone	1-hour	100$^{(4)}$	235$^{(5)}$	235$^{(5)}$
Hydrocarbons	3-hour	100$^{(6)}$	-	-

NOTES:

1. Not to be exceeded more than once per year.
2. Annual geometric mean; all other annual standards are arithmetic means.
3. Guideline only.
4. Daylight hours only. Standard is for photochemical oxidants.
5. Standard attained when expected number of days with maximum hourly average above the standard is equal to or less than one.
6. Morning hours only.

TABLE 5.4-3

PREVENTION OF SIGNIFICANT DETERIORATION INCREMENTS[1]
($\mu g/m^3$)

		Class I	Class II	Class III
Sulfur Dioxide	Annual	2	20	40
	24-hour	5	91	182
	3-hour	25	512	700
Particulates	Annual[2]	5	19	37
	24-hour	10	37	75

NOTES:

1. 3-hour and 24-hour increments may be exceeded once per year.

2. Annual geometric mean.

TABLE 5.4-4

FEDERAL SIGNIFICANT EMISSION RATES[1]
(tons/year)

Sulfur Dioxide	40
Particulates	25
Nitrogen Oxides	40
Carbon Monoxide	100
Ozone (volatile organic compounds)	40
Lead	0.6

NOTE:

1. Hawaii DOH has proposed equivalent level for significant emissions.

TABLE 5.4-5

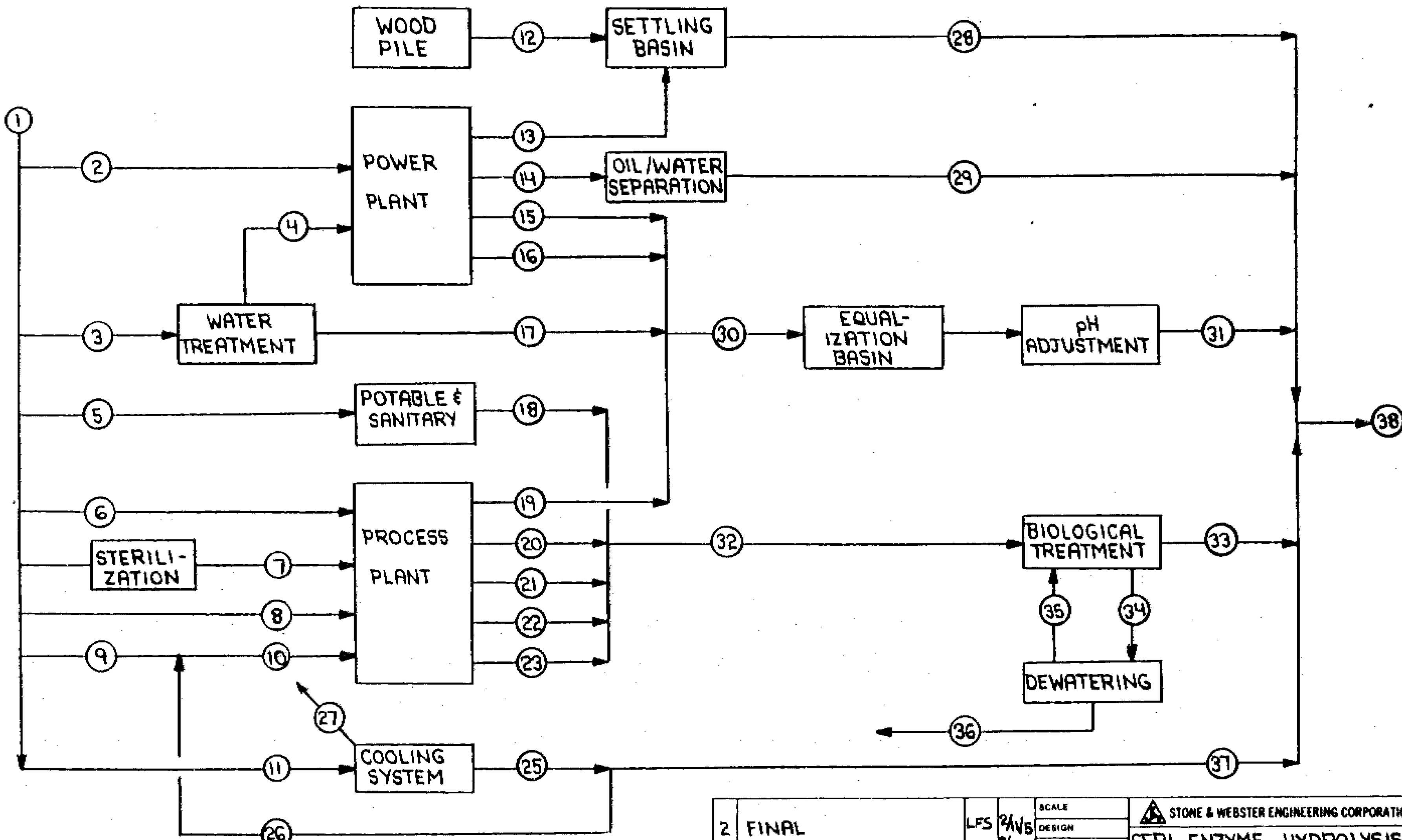
BEST AVAILABLE CONTROL TECHNOLOGY AND AIR EMISSIONS[1]

Pollutant	Control Method	Emission Limit ($lb/10^6$ Btu)	Annual Emissions[2] (ton/yr)
Particulates	Baghouse	0.03	30
Nitrogen Oxides	Fluidized Bed	0.4	300-500[3]
Carbon Monoxide	Combustion Control	-	<100[4]
Volatile Organic Compounds		-	<1[5]
Sulfur Dioxide	-	-	negligible
Lead	-	-	negligible

NOTES:

1. BACT is determined by Agency on a case-by-case basis. Control methods proposed are based on experience with similar facilities.
2. Annual emissions conform to BACT emission limits and are based on pilot plant test data, with the exception of VOC.
3. Test data ranged from 0.3 to 0.5 $lb/10^6$ Btu. Commercial unit should attain 0.4 $lb/10^6$ Btu.
4. Pilot plant not optimized for CO reduction, but commercial unit should attain CO levels comparable to a similarly sized pulverized coal-fired boiler (<100 ton/yr).
5. Pretreatment VOC emissions include process (29.7 ton/yr) and storage and withdrawal (0.3 ton/yr). VOC emissions will be controlled to a 90 percent reduction by combusting the recovery vent stream in the boiler and treating the carbon dioxide wash column vent stream in a carbon adsorber.

UTILITY WATER BALANCE



NO.	REVISION DESCRIPTION	BY	DATE		
2	FINAL	LFS	2/4/85	SCALE	STONE & WEBSTER ENGINEERING CORPORATION
1	FINAL DRAFT	LFS	1/29/85	DESIGN	SERI ENZYME HYDROLYSIS
CERT. FOR FAB.	CERT. FOR CONST.	APPROVED		CHECKED	
				DRAWN LEF	AREA / J.O. NO. 14667 / DRAWING NO. FIG. 5.4-1 / ISSUE 2
				CHECKED	

SECTION 6

PLANT CONSTRUCTION AND OPERATION

6.1 PROJECT CONSTRUCTION SCHEDULE

The project construction schedule for the Hawaiian Enzyme Hydrolysis Facility is shown in Figure 6.1-1. The schedule reflects a 32-month period from project authorization to commercial operation.

The major schedule elements include the following:

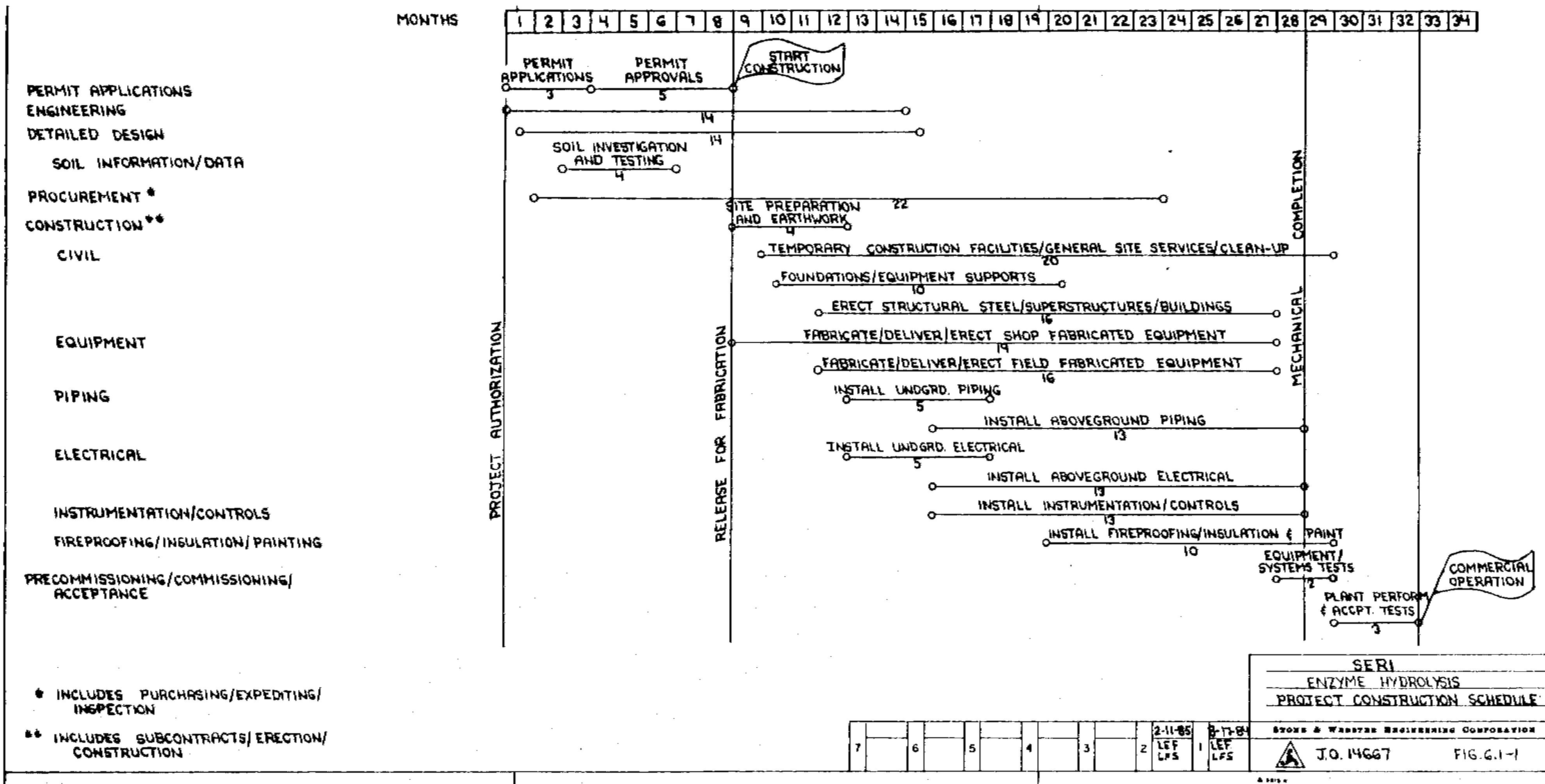
- A 3-month period to prepare and submit permit applications, followed by a 5-month period for agency review and approval process. The end of this period results in a total of 8 months elapsed between project authorization and the start of construction.

- A 14-month period for the Engineering Phase, which commences with the authorization of the project.

- A 14-month period for the Detailed Design Phase, which begins 2 weeks after project authorization.

- An overall 22-month procurement activity, beginning 1 month after the authorization of the project and including a period to initiate procurement activities.

- A 24-month period from start of construction to commercial operation.



MONTHS
1 2 3 4 5 6 7 8 9 10 11 12 13 14 15 16 17 18 19 20 21 22 23 24 25 26 27 28 29 30 31 32 33 34
PERMIT APPLICATIONS
ENGINEERING
DETAILED DESIGN
SOIL INFORMATION/DATA
PROCUREMENT *
CONSTRUCTION **
CIVIL
EQUIPMENT
PIPING
ELECTRICAL
INSTRUMENTATION/CONTROLS
FIREPROOFING/INSULATION/PAINTING
PRECOMMISSIONING/COMMISSIONING/ACCEPTANCE
PERMIT APPLICATIONS 3
PERMIT APPROVALS 5
START CONSTRUCTION
14
14
SOIL INVESTIGATION AND TESTING 4
22
SITE PREPARATION AND EARTHWORK 4
TEMPORARY CONSTRUCTION FACILITIES/GENERAL SITE SERVICES/CLEAN-UP 20
FOUNDATIONS/EQUIPMENT SUPPORTS 10
ERECT STRUCTURAL STEEL/SUPERSTRUCTURES/BUILDINGS 16
FABRICATE/DELIVER/ERECT SHOP FABRICATED EQUIPMENT 19
FABRICATE/DELIVER/ERECT FIELD FABRICATED EQUIPMENT 16
INSTALL UNDGRD. PIPING 5
INSTALL ABOVEGROUND PIPING 13
INSTALL UNDGRD. ELECTRICAL 5
INSTALL ABOVEGROUND ELECTRICAL 13
INSTALL INSTRUMENTATION/CONTROLS 13
INSTALL FIREPROOFING/INSULATION & PAINT 10
EQUIPMENT/SYSTEMS TESTS 2
PLANT PERFORM & ACCPT. TESTS 3
COMMERCIAL OPERATION
PROJECT AUTHORIZATION
RELEASE FOR FABRICATION
MECHANICAL COMPLETION
* INCLUDES PURCHASING/EXPEDITING/INSPECTION
** INCLUDES SUBCONTRACTS/ERECTION/CONSTRUCTION
7 6 5 4 3 2 1
2-11-85 LEF LFS
8-17-84 LEF LFS
SERI
ENZYME HYDROLYSIS
PROJECT CONSTRUCTION SCHEDULE
STONE & WEBSTER ENGINEERING CORPORATION
J.O. 14667
FIG. 6.1-1

6.2 PLANT OPERATION

The enzyme hydrolysis based wood-to-ethanol facility is staffed as outlined in Figure 6.2-1. The overall reponsibility for the facility rests with the Plant Manager. Assisting the Plant Manager are an Operations Superintendent, Plant Engineer, Maintenance Superintendent, and administrative staff. It is assumed in this staffing that the facility is under the jurisdiction of a larger corporation and the administrative functions relating to payroll, personnel, and other corporate affairs are delegated to the home office.

The actual operation of the plant is the responsibility of the Operations Superintendent. Reporting to the Operations Superintendent is the Shift Supervisor who actually oversees the plant operations on each shift. The Shift Supervisor is directly responsible for the Control Room, plant operators, and the material handlers. In addition, in his capacity of directing plant operations, he supervises the Waste/Water Treatment Section and the Quality Control/Microbiological Laboratory.

The Control Room will be the center of operations. Using remote control, many of the plant evolutions will be conducted by the Control Room staff. In addition, plant functions that cannot be conducted directly from the Control Room will be initiated by its operators. The Shift Supervisor and two operators will staff the Control Room to monitor process flows and conditions, ensuring that the process is run properly and efficiently.

Plant operations requiring local attention will be conducted by the plant operations staff. Nine operators will be available to perform remote operations as guided by the Control Room. Additional responsibilities of the plant operators will include assisting in general maintenance and overall plant cleanliness. During the day shift, two additional operators will be present to assist the normal staff in conducting special maintenance or plant evolutions.

Incoming raw materials and chemicals and the distribution of product are also the responsibility of the Shift Supervisor. Four material handlers will be present on each shift, plus an additional two during the day, to maintain the wood pile and oversee the operation of the feedstock handling, chemical storage, and product blending sections of the plant.

The Plant Engineer is responsible for the Waste/Water Treatment and Quality Control/Microbiological Laboratory. This structure is to remove the quality control function from the operating section of the plant, thus ensuring its impartiality.

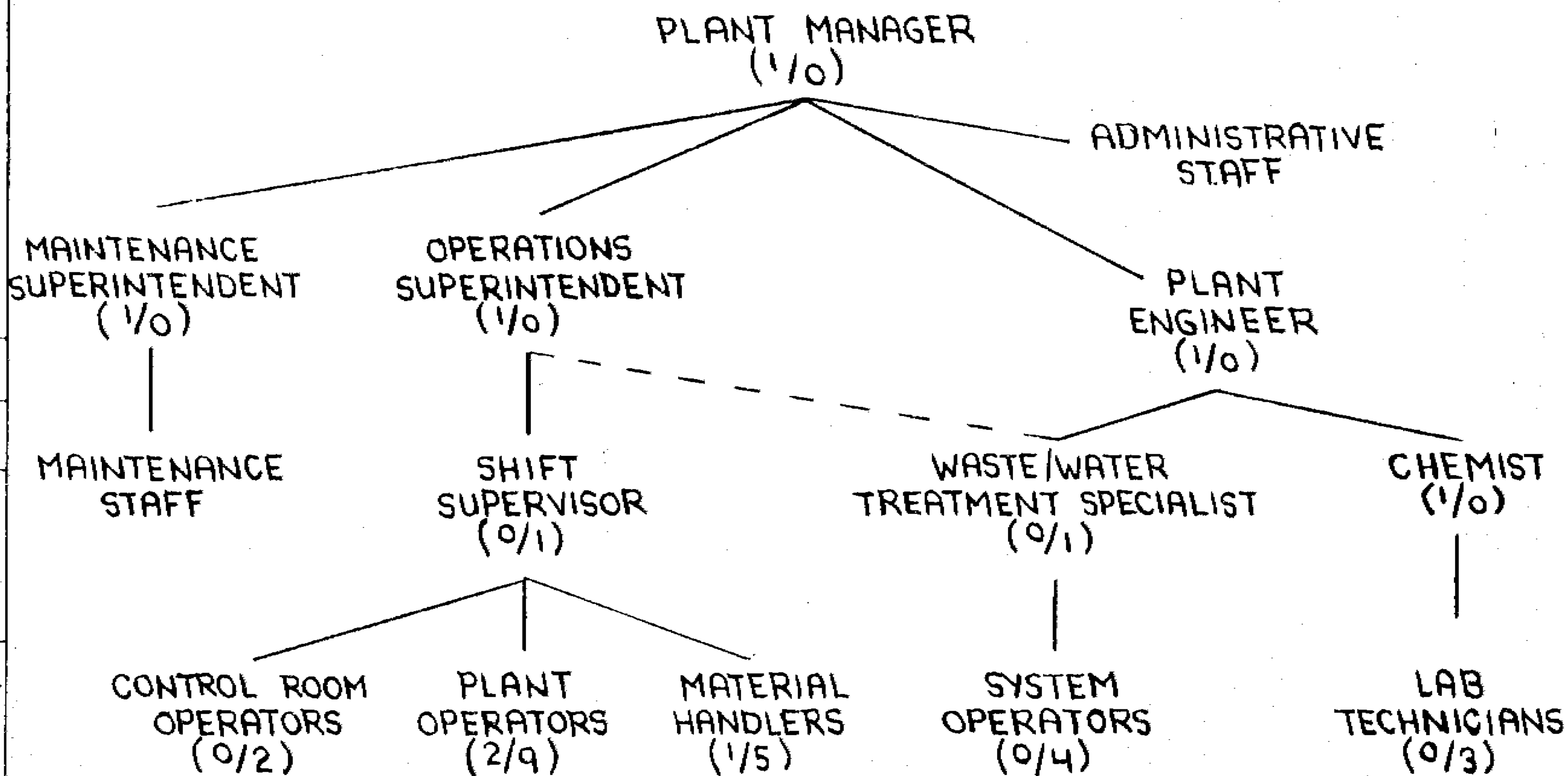
The Waste and Water Treatment Systems (Sections 1200 and 1500, respectively) make up a significant portion of both the capital cost and plant area in this facility. A separate Waste/Water Treatment Specialist will report to the Plant Engineer and be responsbile for the proper operation of these systems. Included in this staff of four system operators per shift will be personnel meeting the licensing requirements in the State of Hawaii for treatment plant operators.

To provide both the raw material and product analyses required for quality control, and the necessary microbiological support, an onsite laboratory is provided and staffed. The laboratory will be manned on all shifts by three qualified technicians. A Chemist/Microbiologist will oversee the daily operation and be responsible for maintaining laboratory analytical results and standards.

Reports concerning the final product quality and the facility effluent discharged will be completed by the Plant Engineer, based upon the reports of his staff.

General plant maintenance is provided by a maintenance staff under the supervision of the Maintenance Superintendent. For the feasibility study cost analysis, maintenance labor was considered as a factor of the installed equipment cost. The actual manning of the maintenance shop is not defined. Typically for a facility of this size, the maintenance staff would include instrument technicians, electricians, mechanics, and a welder. Plant shutdowns or special maintenance requirements would be staffed on a contracted basis.

PLANT STAFFING



TOTAL STAFF = 8/25

LEGEND (A/B) A = PERSONNEL PRESENT ONLY ON DAY SHIFT (MON-FRI)
B = ROTATING PERSONNEL ON EACH SHIFT

NO.	REVISION DESCRIPTION	BY	DATE
2	FINAL	LFS	2/1/75
1	FINAL DRAFT	LFS	8/17/74

DRAWN LEF

STONE & WEBSTER ENGINEERING CORPORATION

SERI - ENZYME

J.O. NO. 14667 — FIG. 6.2.1 — ISSUE 2

SECTION 7

ECONOMIC ANALYSIS

The economic feasibility of the enzymatic-based ethanol plant is presented based on the Nth plant site near Hilo, Hawaii. The analysis includes the economic results of potential improvements via research and development, as well as a sensitivity analysis based on critical parameters.

7.1 SUMMARY

The investment analysis of the enzymatic-based ethanol plant takes into consideration the variables in the operation of the plant, the price of raw materials, and product selling price by performing a sensitivity analysis. The sensitivity analysis considers variations in operational availability (stream factor), raw materials cost (wood, electricity), by-products (furfural, C_5 to ethanol, and lignin), and capital costs. The total facilities investment is estimated to be $150,624,000, including an allowance for indeterminants. The potential increase in the total facilities investment due to the current level of process development should not be ignored.

The sensitivity analysis on the base case is then extended by considering potential technical improvements in the enzyme hydrolysis reactor system, enzyme production system, and other process equipment which affect the plant economics. In addition, changes in the method of financing the initial investment (i.e., the assumption of debt) are investigated.

The analysis is based on a constant dollar basis (1984 dollars), 100 percent equity financing, and a minimum discounted cash flow rate of return (DCFROR) of 15 percent after taxes. The analysis includes the use of the investment tax credit (ITC-10 percent), the energy investment tax credit (EITC-10 percent) and the accelerated cost recovery system (depreciation, ACRS-15, 22, 21, 21, and 21 percent) over 5 years. The ACRS is applied to 90 percent of the depreciable plant in accordance with current regulations. The economic basis is shown in Table 7.1-1. No specific alcohol fuel tax credit is taken since it is assumed that this credit is already reflected in the current selling price for fuel ethanol. The base case plant design utilizes the pentose fraction of the wood and other plant waste streams to produce methane in an anaerobic digestor. The methane is subsequently burned to produce process steam. The wood lignin is also used as boiler fuel. This design results in a required ethanol selling price of $3.50/gal, which is above the estimated ethanol selling price in Hawaii of $1.80-2.00/gal.

A change in the financial basis to include 25 percent debt at a real interest rate of 8 percent reduces the required ethanol selling price to $3.04/gal. This price is still above the estimated Hawaii price of ethanol and indicates that a better use of the pentosan (C_5) fraction of the wood and/or the lignin fraction of the wood is required in order to attract investors to finance this facility.

Two alternative uses for the pentosan wood fraction are the production of ethanol or furfural. The lignin may be sold to the forest products industry as a binder. The potential problems with each route are discussed in Section 8. The potential range of required ethanol selling prices for these two designs (with and without lignin sale) are:

Case	Ethanol Price Range (without lignin sale) ($/gal)	Ethanol Selling Price (with lignin[1] sale) ($/gal)
Pentose to furfural[2]	2.14-3.56	1.42-2.90
Pentose to ethanol	2.08-3.16	1.52-2.65

NOTES:

1. Lignin value at $0.15/lb net to the plant

2. Furfural value at $0.20/lb net to the plant

The details of the selling price range are given in Section 7.6.5.2.

The areas where economic improvements could be achieved through further research and development are increasing enzyme activity and increasing the efficiency of hydrolysis. The recovery of sodium hydroxide also should be included in any future design. The details of these changes are discussed in Section 8. The potential reduction in required ethanol selling price for these goals are:

Component	Reduction in Ethanol Selling Price (¢/gal)
Enzyme Activity	18
Hydrolysis Efficiency	21
Sodium Hydroxide Recovery	15

The above three cases are potentially additive and could have a combined effect of potentially lowering the ethanol selling price by approximately $0.50/gal.

The achievement of these optimistic research goals will not in itself make this process attractive to private investors. The key is in conversion of the pentosan and lignin fractions of the wood to valuable by-products.

TABLE 7.1-1

BASIS FOR INVESTMENT ANALYSIS[1]

Component	Amount
Plant Life, years	20
Operating Factor, hr/yr	8,000
Equity, percent	100
Required Return on Investment, percent	15 DCFROR (after tax)
Interest Rate During Construction, percent	8 real
Ethanol Production Rate, MM gal/yr	15
Tax Rate, percent	50
Investment Tax Credit, percent	10
Renewable Energy Tax Credit, percent	10
Depreciation (on 90 percent of plant)	15, 22, 21, 21, 21
Construction Time, years	2.7
Wood Cost	$39/BDT

NOTE:

1. The investment analysis is based on a discounted cash flow rate of return (DCFROR) in constant 1984 dollars. The cost of capital (Rate of Return) and interest rates are net of inflation.

7.2 ECONOMIC BASIS

The economic feasibility of the enzymatic-based ethanol plant is based on the Nth plant in a series of plants. The major changes in the plants design would be dictated by site-specific conditions, such as climate, environmental requirements, geotechnical limitations, and specific co-products. A discounted cash flow (DCF) type of analysis with a set rate of return to investors of 15 percent is used to determine the ethanol selling price from this plant. To eliminate the need to estimate future inflation rates, the analysis is performed in constant 1984 dollars. The analysis includes the use of the investment tax credit (ITC-10 percent), the energy investment tax credit (EITC-10 percent) and the accelerated cost recovery system (depreciation, ACRS - 15, 22, 21, and 21 percent) over 5 years. The ACRS is applied to 90 percent of the depreciable plant in accordance with current regulations. The economic basis is shown in Table 7.1-1. No specific alcohol fuel tax credit is taken since it is assumed that this credit is already reflected in the current selling price for fuel ethanol. The economic parameters for the plant are given in Table 7.1-1.

7.2.1 Capital Cost Estimation Technique

The base-case conceptual design was costed to allow economic evaluations to be performed. The capital costs for the ethanol plant were determined from vendor quotes, cost data from previous engineering and construction projects, and estimates where direct determination was not possible (see Section 7.3). The plant was assumed to be constructed by contractors using union labor. Labor wage rates are from a recent labor survey generated for the Hilo area.

An Allowance for Indeterminates (AFI) has been included in the capital cost as a percentage of total materials and labor cost (not including subcontractor's fee and overhead), and is based on the known level of detail of the engineering design (e.g., the AFI goes down with a greater level of detail). Each section of the plant was evaluated and an overall AFI determined. The AFI has been set at 18 percent of capital costs ($16,876,000).

The Process Development Allowance (PDA) is a method of recognizing that the actual cost of facilities always exceeds the cost estimated when development is incomplete and the concept is not completely defined. The PDA is applied to process units of the plant that are not presently available on a commercial level. The use of a PDA would allow the capital cost estimate to:

- Account for the increase in the cost of facilities that experience shows always occurs when process development proceeds from one level to a more advanced level, and the definition of the process becomes more detailed.

- Be compared with processes at different stages of development on an equivalent basis.

A PDA was not calculated for this process. It should be recognized that some type of increased capital cost, due to the stage of development and the unique function of several process units, should be added to the base case. This added cost would allow a more valid interpretation of the results.

Prior to financing an actual plant, a risk analysis is typically performed and a separate contingency is calculated. This risk analysis contingency (RAC) is a method of quantifying the probability of cost overruns due to unforeseen events. Typical items included in the RAC would be:

- Inadequacies in plant scoping (e.g., design changes, revised regulations, environmental impacts)
- Insufficient information (e.g., land costs, site preparation)
- Labor uncertainties (e.g., strikes, productivity, contracts)
- Unforeseen materials availability, subcontractor expenses or engineering expenses.

Since exact site-specific information and definitive design details were not fully developed for this plant, the RAC was also excluded from the plant capital costs.

7.2.2 Economic Analysis Methods

The investment analysis takes into consideration the variables in the operation of the plant, the price of raw materials, and product selling price by performing a sensitivity analysis. The sensitivity analysis considers variation in operational availability (stream factor), raw materials cost (wood, electricity), by-product credits, and capital costs.

The investment analysis on the base-case design is then extended by considering technical modifications of the design which improve the economics. In addition, changes in the method of financing the initial investment (i.e., the assumption of debt) are investigated.

7.3 CAPITAL COSTS

The capital costs of the enzymatic-based ethanol plant are shown in Table 7.3-1. The total cost is $150,624,000. A sectional breakdown of these costs is shown in Table 7.3-2. The costs have been divided into materials and equipment, field installation labor, freight and tariffs, engineering and construction management, and working capital. The working capital is based upon 14 days storage of both raw materials and finished product, plus 0.9 percent of the total base cost for spare parts and miscellaneous materials. The basis for the cost estimate is the base-case engineering design performed as a part of this study. A major equipment list (Appendix C) was developed and equipment classified according to the type of cost estimation required. All major equipment was sized and equipment duty specification sheets prepared. The items requiring vendor quotations were assembled and sent for budgetary type estimate quotations. A list of vendors contacted is included in Appendix A. The vendor responses were evaluated and the quotations adjusted to include total equipment costs. Since budgetary type costing information is not indicative of competitive bidding situations, specific vendors were not selected. The vendor quotations were compared with inhouse information to determine equipment costs, delivery schedules, and installation factors for each piece of equipment or vendor package.

This cost estimate is based on the installation of the Nth plant in a series of plants. The Nth plant is assumed to be constructed over a 32-month schedule. The interest during construction is calculated using an S curve (for the expenditure of funds) at a real interest rate (net of inflation) of 8 percent.

TABLE 7.3-1

ENZYME-BASED ETHANOL PLANT
CAPITAL COST - BASE CASE

Component	Cost (1984 dollars)
Material and Equipment	54,885,000
Labor	32,915,000
Freight and Tariffs	2,400,000
Total Base Cost	90,200,000
Land Cost	72,000
Engineering and Construction Management	13,312,000
Allowance for Indeterminents	16,876,000
Total Installed Cost	120,460,000
Initial Catalysts and Chemicals	439,000
Startup Expenses (3 months)	12,068,000
Interest During Construction	14,375,000
Working Capital	3,282,000
Total Facilities Investment	150,624,000

TABLE 7.3-2

SECTIONAL COST FOR ENZYME HYDROLYSIS

Section	Cost
100 - Pretreatment	$2,864,000
200 - Steam Explosion/Wash	$9,691,000
300 - Enzyme Production	$7,242,000
400 - Hydrolysis	$8,186,000
500 - Evaporation	$2,807,000
600 - Fermentation	$4,274,000
700 - Distillation	$3,468,000
800 - Anaerobic Digestion	$7,291,000
900 - Boiler	$15,331,000
1000 - Feedstock Handling	$7,298,000
1100 - Cooling Water	$1,136,000
1200 - Wastetreatment/Vent Scrubbing	$8,297,000
1300 - Chemical Handling	$195,000
1400 - Product Storage and Unloading	$855,000
1500 - Instrument Air/Fire Protection (buildings and miscellaneous auxiliaries)	$8,865,000
	$87,800,000

7.4 REVENUES

The revenues from the ethanol plant come from several sources: the sale of ethanol, the sale of carbon dioxide, the sale of C_5 stream derived by-products, and the potential sale of lignin. In general, raw carbon dioxide can be sold to a distributor for purification and sale for approximately $10/ton. Although there is not a sufficient market in Hawaii to justify the sale of this product, the effect of this revenue on plant economics for other sites is discussed in Section 7.6. The C_5 stream can be converted into several different by-products. The base case assumes the anaerobic digestion of this stream to produce methane-rich gas which is then burned in the lignin boiler to produce steam for in-plant use. Other uses for this C_5 stream include:

- Direct sale of the methane-rich digestion gases to an industrial customer
- The production and sale of animal feed
- The production and sale of furfural
- The production of additional ethanol

A potential alternative is the sale of this by-product (e.g., to the forest products industry for use as an adhesive). Since a definitive market for lignin is not defined, the price of this by-product is varied to establish the potential reduction in ethanol selling price. For similar reasons, the selling price of furfural was also varied to determine the potential reduction in ethanol selling price.

The selling prices of methane and animal feed are set by the local market conditions in Hawaii and are not varied in the investment analysis.

7.5 OPERATING AND MAINTENANCE COSTS

The operating costs for the plant include wood, electricity, chemicals, manpower, water, and waste disposal costs. Tables 7.5-1 to 7.5-4 enumerate these various plant costs. The manpower costs include 100 percent of wages for overhead. The overhead costs follow the general criteria for synfuel plants, as set forth by the Gas Research Institute.

Maintenance costs are calculated as a percentage of plant section base cost. Table 7.5-5 gives the maintenance factor by plant section. Insurance and local taxes are taken as 1.5 percent of the plant total installed cost. Table 7.5-6 summarizes the annual operating and maintenance costs.

TABLE 7.5-1

MAJOR PLANT AND RAW MATERIAL OPERATING COSTS

Component	Unit Cost	Amount (per year)	Costs ($)
Wood	$39/ton (dry)	217,848 ton	8,496,072
Gasoline	$1.00/gal	714,304 gal	714,304
Sulfuric Acid	$92.32/ton	7,635 ton	704,863
Ammonium Hydroxide	$200/ton	4,703 ton	940,600
Sodium Hydroxide	$500/ton	5,772 ton	2,886,000
Wood Fuel	$39/ton (dry)	13,898 ton	542,022

TABLE 7.5-2

CHEMICAL USAGE

Component	Unit Cost	Quantity (Per Year)	Cost ($/yr)
Process Chemicals			
Corn Steep Liquor	$227.00/ton	1,648 ton	374,100
Offsite Chemicals			
100% Sulfuric Acid	$92.32/ton	119 ton	110,000
Chlorine	$550.00/ton	6.5 ton	3,600
50% Sodium Hydroxide	$250.00/ton	281 ton	70,200
29% Ammonium Hydroxide	$200.00/ton	1 ton	200
35% Hydrazine	$2.50/lb	333 lb	800
85% Phosphoric Acid	$1,000.00/ton	168 ton	168,000
Anhydrous Ammonia	$600.00/ton	254 ton	152,400
Liquid Polymer	$4,000.00/ton	16 ton	64,000
Corrosion Inhibitor	$1.83/lb	7,335 lb	13,500
			582,700
TOTAL			956,800

TABLE 7.5-3

MISCELLANEOUS OPERATING COSTS

Component	Unit Cost	Quantity (Per Year)	Cost ($)
Water	$1.25/M gal	408,480 M/gal	510,600
Sludge Disposal	$10/ton	16,340 ton	163,400
Fermentation Expenses	-	-	54,600
Operating Supplies	-	-	120,500
			849,100

TABLE 7.5-4

OPERATING LABOR SUMMARY

	No. of Men/Day
Plant Operating Personnel	57
Control Room Personnel	6
Lab Personnel	10
Total	73

73 men/day at $10/hr	$1,927,200/yr
Supervision at 25%	$ 481,800/yr
Total labor	$2,409,000/yr
Overheads	
General (45%)	$1,084,050/yr
Corporate (30%)	$ 722,700/yr
Benefits (25%)	$ 602,250/yr
Total labor including overheads	$4,818,000/yr

TABLE 7.5-5

MAINTENANCE FACTOR BY PLANT SECTION

Plant Section		Maintenance Factor (% of Capital Cost)
100	Pretreatment	6
200	Steam Explosion	5
300	Enzyme Production	5
400	Hydrolysis	5
500	Evaporation	3
600	Fermentation	4
700	Distillation	3
800	Anaerobic Digestion	3
900	Boiler	4
1000	Feedstock Handling	6
1100	Cooling Water	2
1200	Waste Treatment/Vent Scrubbing	6
1300	Chemical Handling	3
1400	Product Storage and Unloading	2
1500	Offsites	2 1/2

TABLE 7.5-6

SUMMARY OF OPERATING COSTS
BASE CASE (1)

Component	Price ($/yr)	Ethanol Cost Contribution (¢/gal)
Raw Materials		
Wood to Process	8,496,072	56.6
Sulfuric Acid	704,900	4.7
Sodium Hydroxide	2,886,000	19.2
Ammonia	940,600	6.3
Wood to Boiler	542,022	3.6
Gasoline	714,304	4.8
Process Chemicals	374,100	2.5
Offsite Chemicals	582,700	3.9
Operating Labor	4,818,000	32.1
Maintenance (Labor & Supplies)	3,660,237	24.4
Electricity	4,644,000	31.0
Insurance and Taxes	1,768,900	11.8
Miscellaneous	849,100	5.7
	30,980,935	206.7

NOTE:

1. 15×10^6 gal/yr ethanol production

7.6 INVESTMENT ANALYSIS

The basis for the economic analysis is presented in Section 7.2. The base-case design utilizes the pentosan (C_5) stream to produce methane which is burned with the lignin in a boiler to produce process steam. The alternative uses of the pentosan stream and the potential sale of the lignin are discussed in Section 7.6.3. The sale of by-products is set at the prevailing price level, where possible. Uncertainty in the by-product price structure was eliminated by establishing a reasonable range of potential prices.

The analysis determined the required selling price of ethanol to meet an assumed minimum rate of return that would be acceptable to an investor.

The minimum rate of return that would be acceptable to an investor is dependent upon the level of risk that is perceived. Two components of risk are apparent in this project:

- Inability to sell the ethanol product and by-products
- Potential problems in operation

The inability to sell the ethanol product and by-products is of prime concern; while the technological problems would be known from the pilot plant and operation of previous plants. The risk in selling the product is concerned with market penetration. In the context of the constant dollar nature of this analysis and the present level of real rate of return obtainable on secure debt securities (approximately 8 percent), an after-tax DCFROR of 15 percent was assumed as the minimum rate of return that an investor would expect for this project.

7.6.1 Ethanol Selling Price

The variation of the required ethanol selling price to satisfy various DCFROR values is shown in Figure 7.6-1. The required ethanol selling price for the base case (C_5 sugars and lignin to process steam) to achieve the desired DCFROR of 15 percent is \$3.50/gal. The current (8/15/84) posted prices of fuel grade ethanol range from \$1.47/gal in Iowa to \$1.67/gal in California. It can be assumed that with transportation costs, ethanol prices in Hawaii would be in the \$1.80-2.00/gal range.

A comparison of the required selling price and the potential selling price indicates that better economic use of the pentose (C_5) sugar and the lignin stream is required to interest investors in this process. The other methods of pentose sugar utilization are compared in Section 7.6.3, along with the economic potential from the sale of lignin.

The relaxation of the requirement of 100-percent equity will also reduce the required ethanol selling price. The use of 25 percent debt at a real interest rate of 8 percent results in a required ethanol selling price of $3.04/gal. The assumption of debt is discussed in Section 7.6.4.

7.6.2 Sensitivity Analysis

The change in the required ethanol selling price with variations in parameters, such as capital cost, raw material prices, plant operation and by-product credits (for the base case) requires a sensitivity analysis to be performed. This analysis will establish a range of ethanol selling prices for the DCFROR of 15 percent.

7.6.2.1 Capital Costs

Figure 7.6-2 shows the effect of a change in capital cost on the required ethanol selling price. A variation in capital cost of approximately 15 percent ($22 million), changes the required ethanol selling price by $0.20/gal. The base-case economics assumes the design and construction of an Nth plant, but no specific process development allowance has been applied to the base-case capital cost. As stated in Section 7.2, an increase in capital costs is anticipated due to the present state of development of this process.

7.6.2.2 Stream Factor

The sensitivity of ethanol selling price to the on-line availability of the plant (stream factor) to produce the ethanol product and by-products is shown in Figure 7.6-3. The base case assumes that the plant operates 8000 hours per year or 91 percent of the available time. A reduction in availability to just 80 percent results in an increase in required ethanol selling price of $0.26/gal. This calculation demonstrates the critical nature of the availability and operability of the plant on the required ethanol selling price.

7.6.2.3 Raw Material Costs and By-product Credits

The sensitivities of raw material prices (i.e., wood, electricity, process chemicals, and of base case by-product revenue, e.g., carbon dioxide) can be analyzed as direct changes in the ethanol selling price. Any change in net revenues of $750,000 is equivalent to $0.05/gal change in ethanol price. Table 7-6-1 gives the results of a unit price change in each of these commodities. Figures 7.6-4 to 7.6-6 show the ethanol price sensitivity to changes in the price of wood, electricity, and carbon dioxide. Of these, wood exhibits the largest effect.

7.6.3 Alternative Uses of Pentose (C_5) Sugars

The need to produce more revenue from the pentose sugars, in order to reduce the required ethanol selling price, has led to the investigation of potential alternatives. Each final product is associated with a different unknown. The unknowns range from the potential value associated with furfural and methane to the ultimate saleability of animal feed. The product revenues are based on costs at the plant site.

Actual selling prices would be higher than this price. The potential products of the pentose sugars are:

- Methane for Direct Sale

  The base-case plant produces 66.9 MM Btu/hr of a methane-rich gas (60 percent CH_4, 40 percent CO_2) which is burned in the boiler to produce steam for internal plant usage. An alternative is to sell this methane-rich gas to an industrial user and import wood to fire the boiler. On the island of Hawaii, the cost of gaseous fuel for industrial use is $11.90/MM Btu. If it is assumed that the methane-rich gas can be sold for $10/MM Btu to the customer ($1.90 is assumed for cleanup and transportation), the required ethanol revenue is reduced by $0.23/gal. This high level of methane revenue is only applicable to the current site. The current price of nonregulated industrial gas in the continental U.S. is approximately $3.50/MM Btu. This reduction of $0.23/gal in the required ethanol selling price is not sufficient to sell the ethanol competitively in Hawaii.

- Animal Feed

  The production of a sugar-based (molasses type) feed for animals (e.g., cattle) is a potential alternative. The feed would have to be tested to determine its acceptability to the cattle. Presently the market in Hawaii has a surplus of molasses and the potential price is approximately $44/ton. In the continental U.S., this low protein feed is worth less than $80/ton. The result of selling this product and importing additional wood is to decrease the required ethanol selling price by $0.12/gal (in Hawaii). This alternative is not sufficient to sell ethanol competitively.

- Additional Ethanol

  The fermentation of the recovered pentose sugars to ethanol results in a reduction in the ethanol selling price by $0.71/gal. The technical details of this approach are discussed in Section 8. The production of additional ethanol can be readily sold with the primary ethanol product and does not entail the complications associated with selling a second product. This type of pentose sugar utilization would be preferred if the ultimate economics would be positive.

- Furfural Production

  Production of acetic acid and furfural from the pentose fraction of the wood is discussed in Section 8. The principal question to address when considering furfural production is the ability to sell the product. The 1982 U.S. production of furfural was 140 million pounds, a single 15 MM gal/hr ethanol plant could produce approximately 38 million pounds. It is obvious that additional markets for this product must be developed in order to justify the saleability of the product. A recent report by SERI, "The Value of Furfural/Ethanol Coproduction from Acid Hydrolysis Processes", indicates that at sufficiently low costs of furfural ($0.20-$0.30/lb), large additional markets for this chemical will develop for substitution in current processes and products.

The effect of the combined production of furfural and acetic acid on the required selling price of ethanol depends on the projected selling price of furfural. The sensitivity is shown below.

Furfural Price ($/lb)	Reduction in Ethanol Selling Price (¢/gal)
0.10	20
0.20	45
0.30	71

Approximately $0.07/gal of the reduction is due to the sale of the acetic acid (at $0.27/lb).

- Lignin

The sale of lignin is dependent on the type and quality of lignin and the overall market for this product. Currently, sodium lignate is sold commercially as a product of pulp production. The similarity of this sodium lignate to the lignin by-product from the sodium-lignin wash in this plant is discussed in Section 8. The potential market and price of this type of lignin is undefined.

The effect of lignin production and sales on the required selling price of ethanol depends on the projected selling price for the lignin. The sensitivity is shown below:

Lignin Selling Price (¢/lb)	Reduction in Ethanol Selling Price (¢/gal)
10	42
15	69
20	97

7.6.4 Debt Financing

The base case analysis is based on the assumption of total equity financing for the ethanol plant. The investment in the plant can be partially financed by debt. A reasonable level of debt for this type of facility to be financed by a long-term (20-yr) bond would be 25 percent. The interest rate on the debt is based on a zero inflation rate, since the economic calculations are performed in constant 1984 dollars. The historic real rate of return (net of inflation) on secured debt is approximately 4 percent, although the current rate is near 8 percent. The reduction in the required ethanol selling price for these two cases are:

Equity	Debt	Net Interest Rate	Reduction in Ethanol Price
75%	25%	4%	\$0.57/gal
75%	25%	8%	\$0.46/gal

7.6.5 Discussion of Results

7.6.5.1 Effects of Individual Sensitivity Parameters

The base-case analysis, at 100-percent equity financing, requires an ethanol price of \$3.50/gal to satisfy the assumption of a 15 percent DCFROR. The inclusion of 25 percent debt (at 8 percent) reduces this required price to \$3.04/gal. This price is above a reasonable selling price for fuel grade ethanol and indicates that more revenue is required from the pentose and lignin fraction of the wood. The effect of the revenues, from the several pentose conversion schemes with and without the sale of lignin, on ethanol selling price are:

Case	Ethanol Selling Price[1] w/Lignin Sales (\$/gal)	w/o Lignin Sales (\$/gal)
Methane	2.12	2.81
Animal Feed	2.23	2.92
Ethanol	1.90	2.43
Furfural (at \$0.20/lb)[2]	1.90	2.59

NOTES:

1. With 25 percent debt at 8 percent interest, lignin sold at \$0.15/lb net to the plant.

2. Net price to the facility, including the sale of acetic acid at \$0.27/lb.

For the base-case plant configuration with the stipulated financial assumptions, the sale of the pentose and lignin fraction of the wood at a relatively high net price to the facility is required for this approach to be attractive to private investors.

7.6.5.2 Combination of Sensitivity Parameters

The sensitivity analysis has considered the effect of individual parameters on the economics of the plant. The question arises as to the effect of a combination of parameters of the site-specific case of Hawaii. Two cases are considered, a pessimistic scenario and an optimistic scenario. Since the base-case analysis has shown that more revenue is required from the pentose sugars and lignin, the combination of sensitivity parameters will concentrate on additional ethanol production or, furfural production along with the sale of lignin. The optimistic scenario considers a 15 percent decrease in capital costs, the sale of carbon dioxide at $10/ton, a decrease in wood price of $10/dry ton, the assumption of debt, and an increase in the stream factor to 95 percent. The pessimistic scenario considers a capital cost increase of 25 percent, an increase in the wood price of $10/dry ton, the assumption of debt and a decrease in stream factor to 70 percent.

The furfural case assumes a net furfural price to the plant of $0.20/lb. The lignin is sold at a net price to the plant at $0.15/lb. The ranges of required ethanol selling price between optimistic and pessimistic scenarios for both cases of pentose conversions with and without lignin sale are:

Case	Required Ethanol Selling Price w/Lignin Sale ($/gal)	Required Ethanol Selling Price w/o Lignin Sale ($/gal)
Pentose to Ethanol	1.52-2.65	2.08-3.16
Pentose to Furfural	1.42-2.90	2.14-3.56

The need to sell the lignin by-product or reduce the base-case production costs is demonstrated in the optimistic case calculations. Selling the ethanol or the furfural (produced from the C_5 fraction) along with lignin results in a favorable ethanol selling price, under optimistic circumstances. The furfural case shows a slightly lower ethanol selling price, but the current market for this product would be met by approximately three commercial facilities of the current design and scale. The development of an expanded furfural market as a result of low priced furfural is open to question.

TABLE 7.6-1

ETHANOL SELLING PRICE CHANGES DUE TO COMMODITY PRICE CHANGES
(Base Case - Pentose Sugars to Steam)

Commodity	Unit Price Change	Equivalent Ethanol Price Change (¢/gal)
Wood	+$10/dry ton	+15
Carbon Dioxide	+$10/ton	-3
Electricity	+1¢/kWh	+4.5
Sodium Hydroxide	+$50/ton	+2.0

FIGURE 7.6-1

ETHANOL PRICE VS. DCFROR

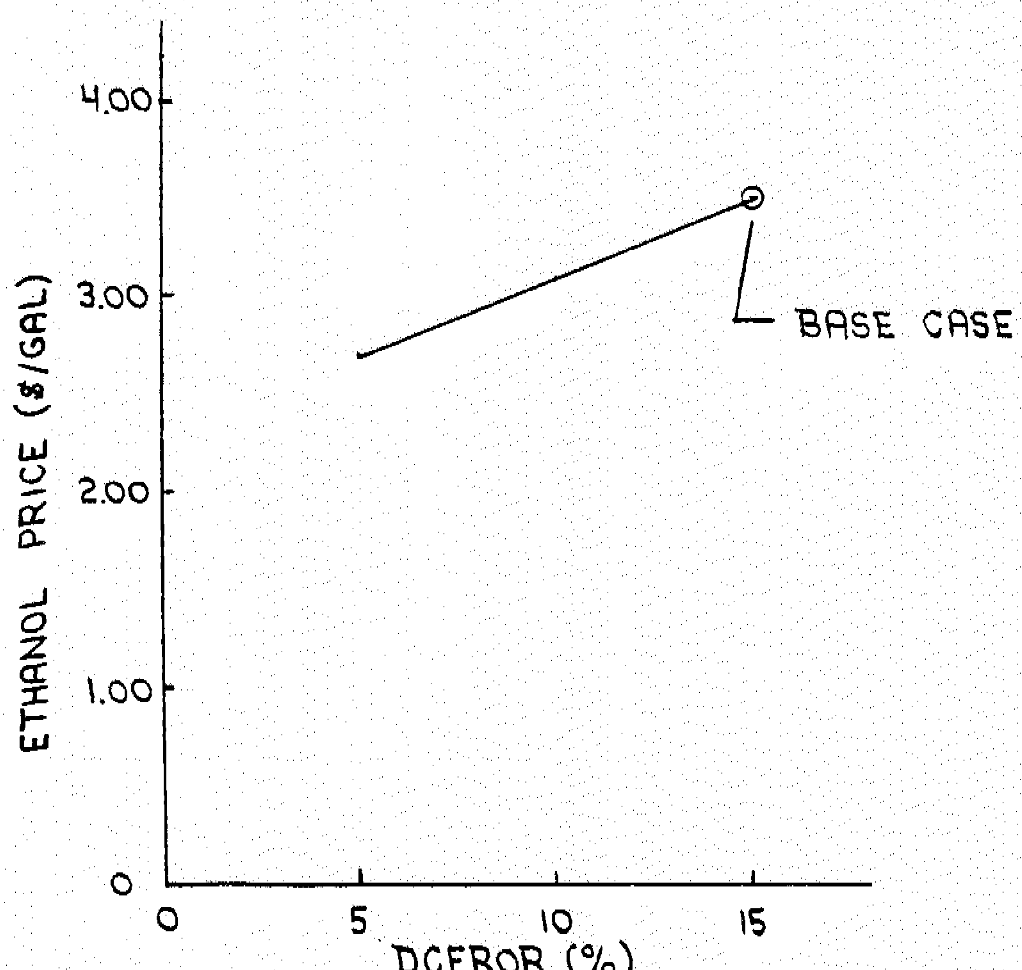


FIGURE 7.6-2

ETHANOL PRICE VS. CAPITAL COST

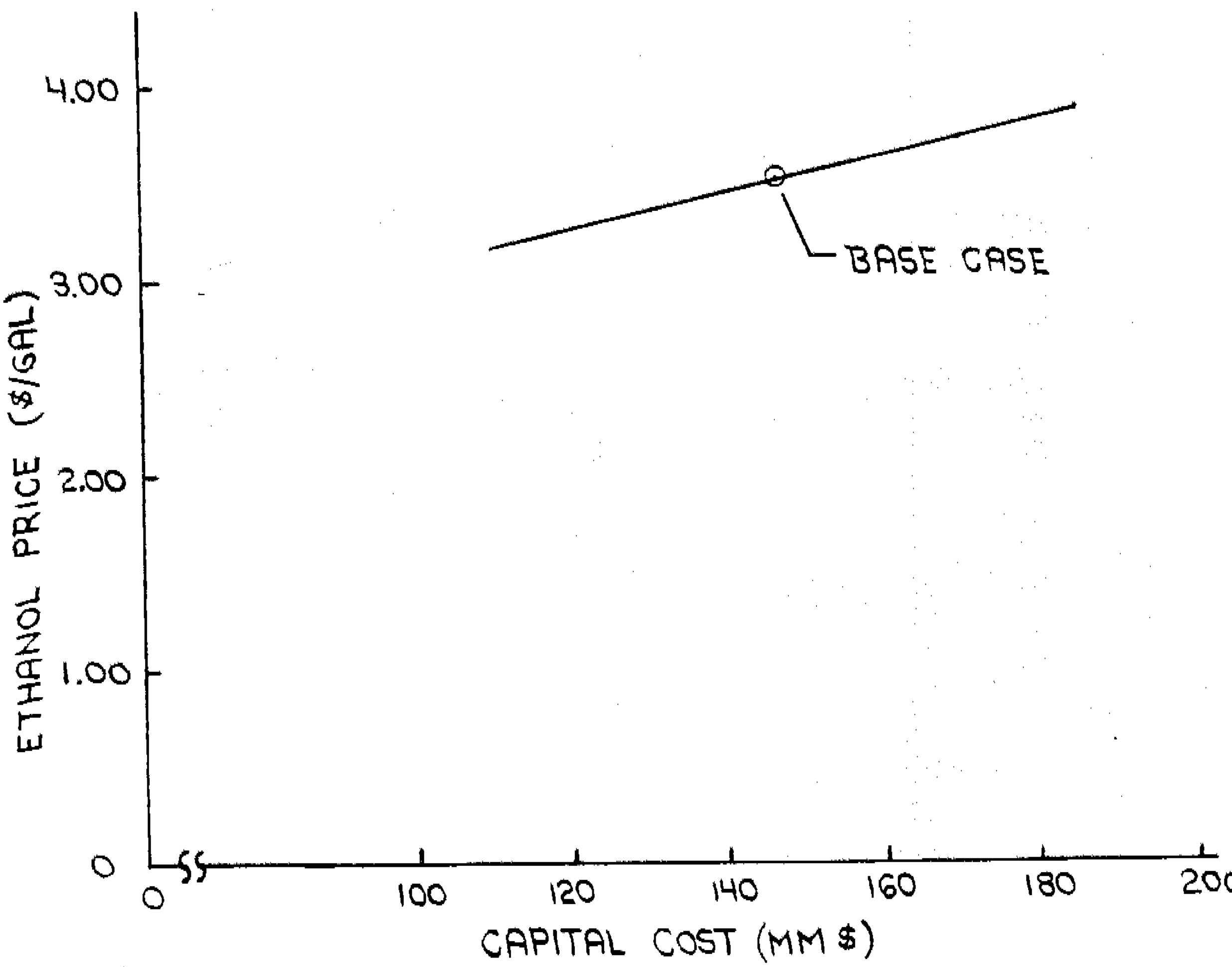


FIGURE 7.6-3

ETHANOL PRICE VS. STREAM FACTOR

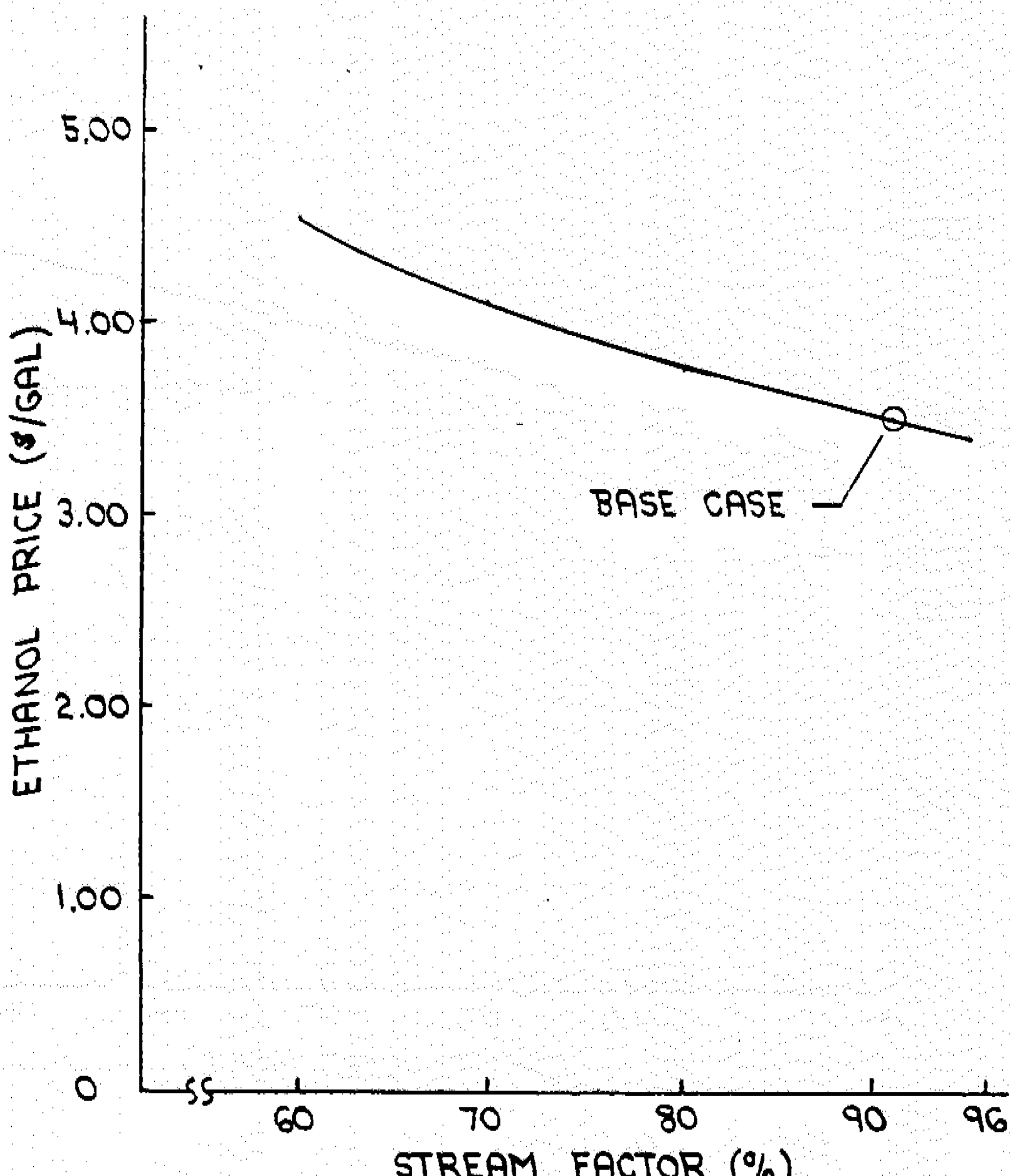


FIGURE 7.6-4

ETHANOL PRICE VS. WOOD PRICE

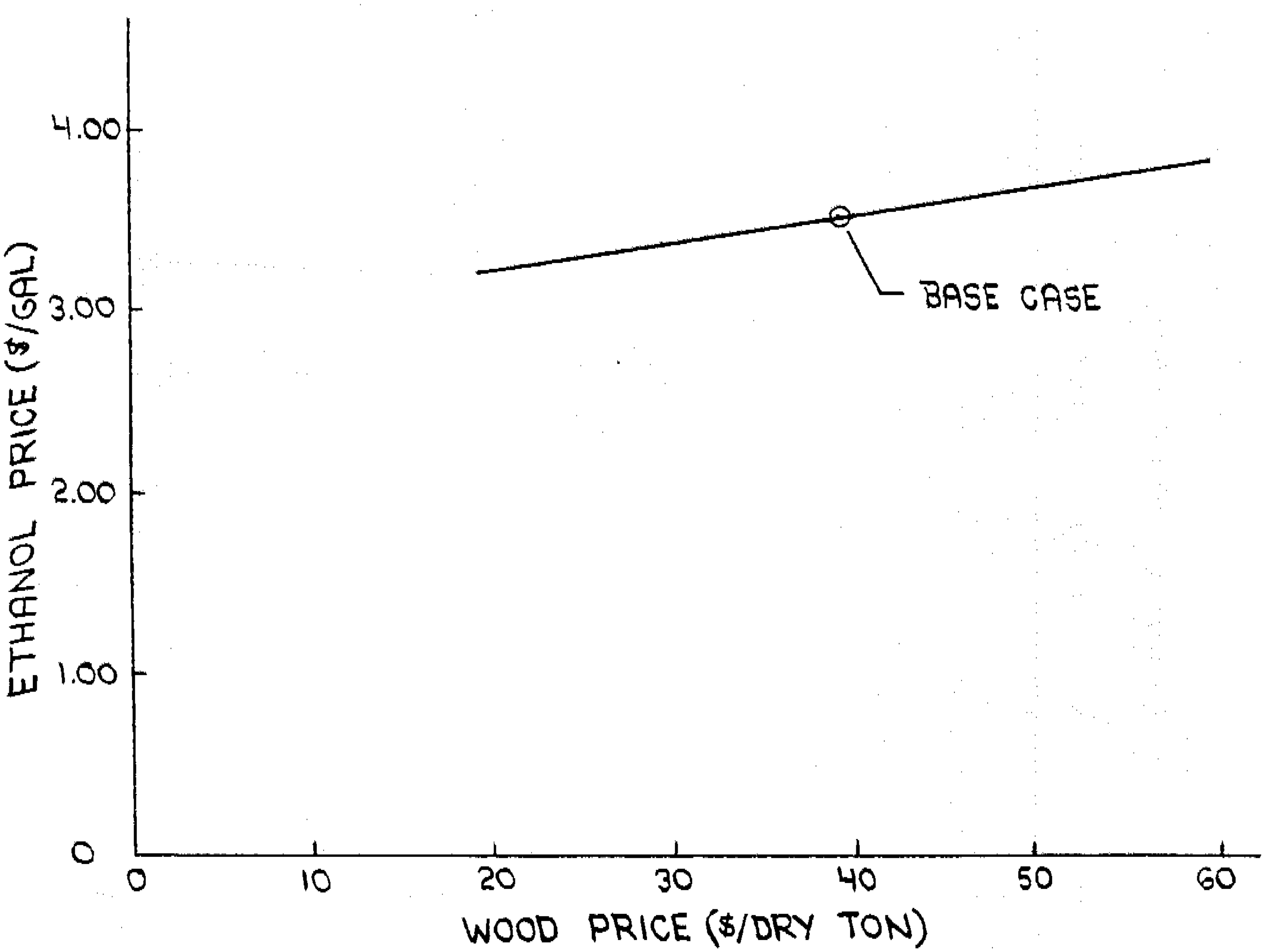


FIGURE 7.6-5

ETHANOL PRICE VS. COST OF ELECTRICITY

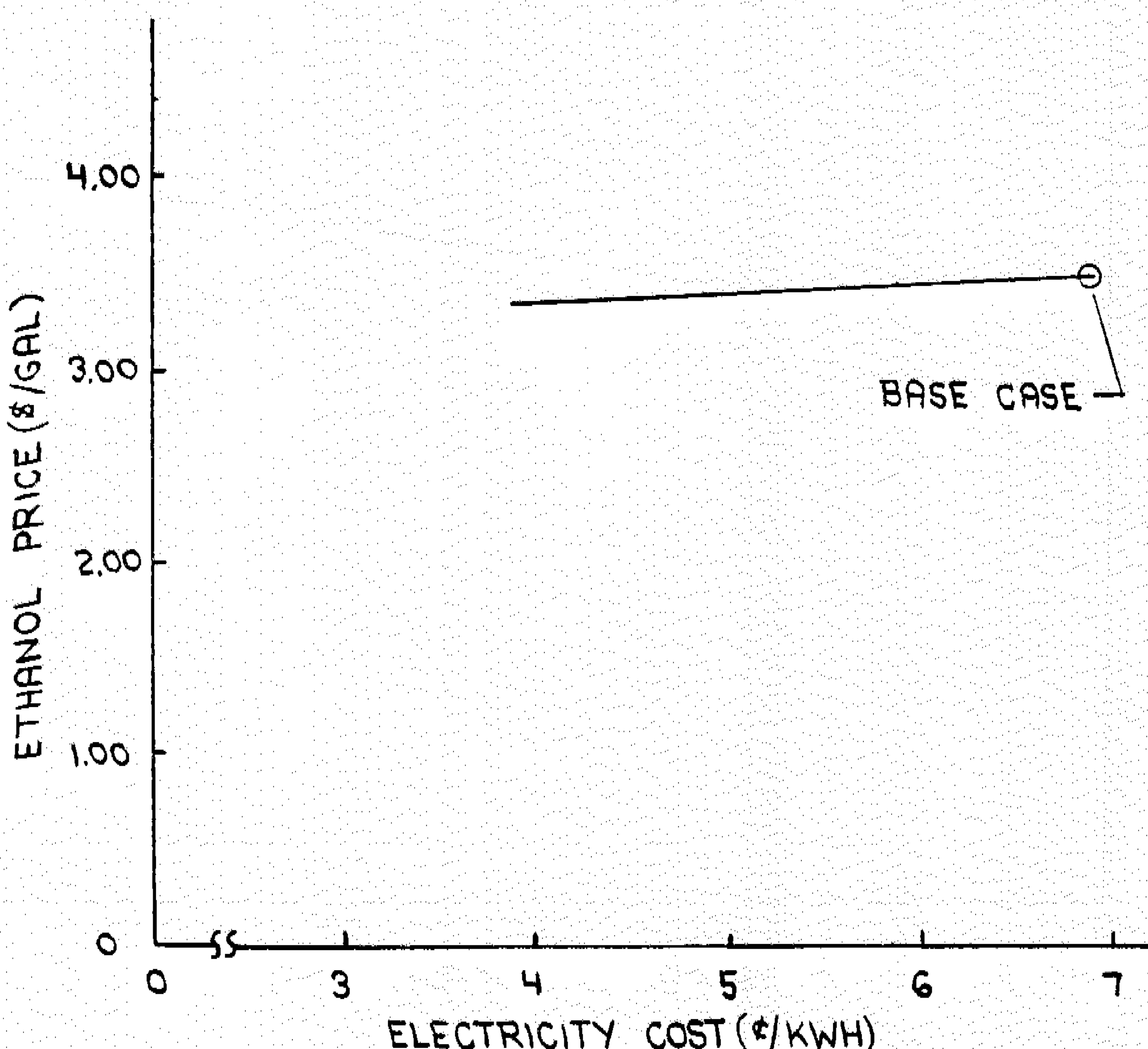
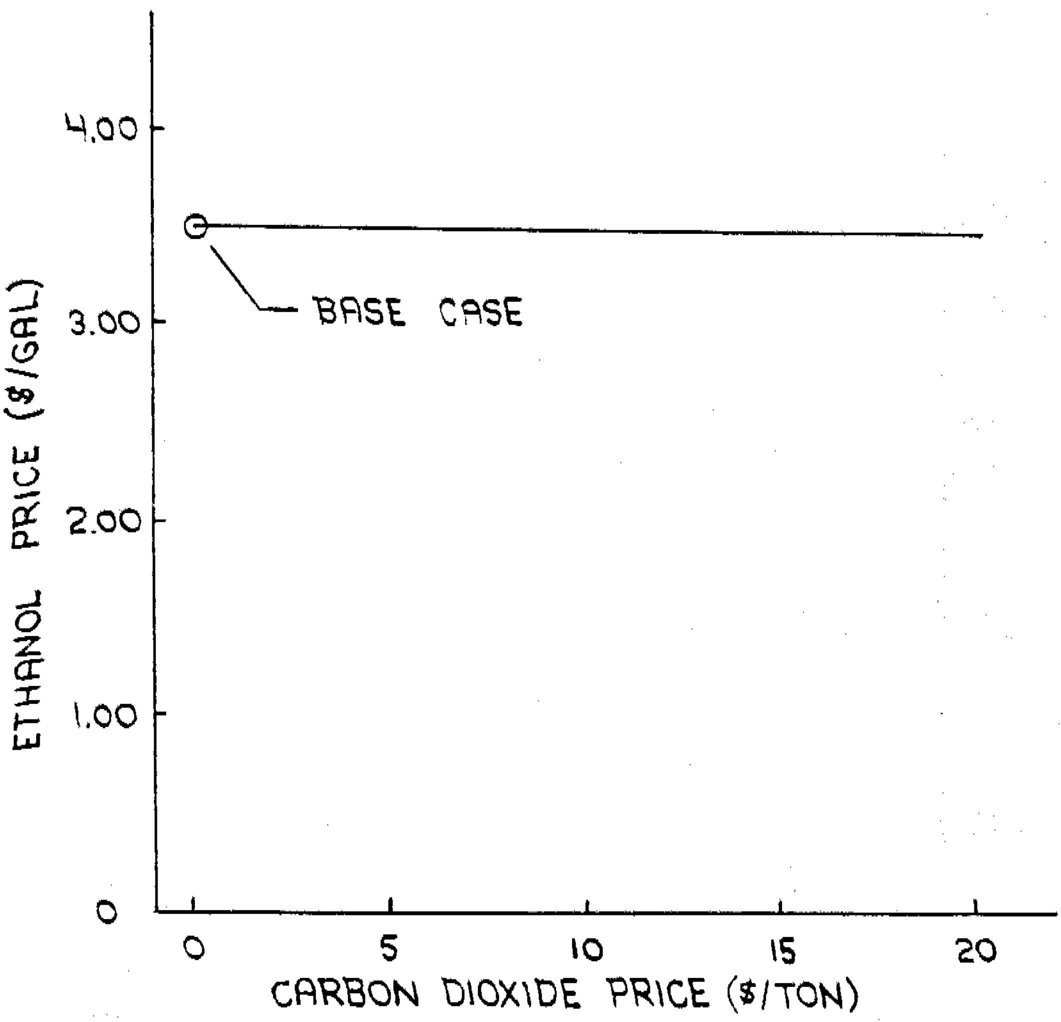


FIGURE 7.6-6

ETHANOL PRICE VS. CARBON DIOXIDE PRICE



7.7 ALTERNATIVE DESIGNS

The analysis of the use of the pentose sugars and sale of lignin shows that, under optimistic conditions, the required ethanol selling price could become competitive with ethanol selling price for both ethanol and furfural production with lignin sale. Since the marketability of either lignin or furfural at the suggested price is uncertain, improvements in the base-case design were investigated for potential economic advantages.

Two areas were investigated to determine their effect on the base-case enzyme production and hydrolysis.

- Hydrolysis reactor yield, residence time and solids concentration
- Enzyme activity and enzyme production residence time

In addition, six design changes on commercial equipment were investigated to determine the economic changes:

- Elimination of evaporation
- Defribration as opposed to steam explosion
- Sodium hydroxide recovery
- Molecular sieve dehydration
- Changes in the chip acid soak
- Lignin separation after hydrolysis

The technical details of these alternative designs are discussed in more detail in Section 8. The elimination of the evaporator may cause difficulties in the fermentation process, and removing the lignin after hydrolysis may render this product non-saleable without expensive treatment. The other changes are conceptually benign. The effect of the trade-off changes to the process are shown below for the above cases:

Component	Reduction in Required Ethanol Selling Price (¢/gal)
Increase in enzyme activity	18
Enzyme residence time reduction	9
Hydrolysis yield, residence time, and concentration improvements	21
Evaporation eliminated	6

Component	Reduction in Required Ethanol Selling Price
Defibration substitution	7[1]
Sodium hydroxide recovery	15
Elimination of chip soak	4
Molecular sieve substitution	3
Lignin separation after hydrolysis	21

[1] Increase in Ethanol Selling Price

Significant improvements in process economics via changes in plant design can occur by limiting the net usage of sodium hydroxide. This is accomplished by either sodium hydroxide recovery in the lignin recovery before hydrolysis or elimination of sodium hydroxide use by separation of the lignin after hydrolysis. The sodium hydroxide recovery is economically advantageous, since it retains the potential for sale of the lignin without additional cost.

Potential improvements in the process are in the areas of enzyme production and hydrolysis. A combination of these technologies has the potential of reducing the ethanol selling price by approximately $0.35/gal. Both of these could be considered research goals.

The economic advantage of the maximum combination of these process improvements ($0.55/gal) still requires conversion of the C_5 sugars and/or the lignin to saleable by-products.

Even if several of the design alternatives could be implemented in combination, the net effect would be only about $0.55/gal reduction in the required ethanol selling price. Conversion of the pentose sugars and sale of lignin are still required to obtain a viable selling price.

7.8 COMPARISON OF SITES

The site selected for the ethanol plant is near Hilo, Hawaii. The background on this site selection is given in Section 3. An investigation was conducted to identify the potential changes in the process economics by siting this plant in a continental U.S. location. The potential advantages in moving the plant are associated with reduced transportation costs for certain chemical raw materials. In addition, the location of the plant in a heavily forested area with an existing wood harvesting infrastructures may reduce feedstock raw material costs.

The comparison site selected was Spokane, Washington. This location was the site for a recent SERI wood gasification to methanol study performed by Stone & Webster Engineering Corporation and has the attributes of an abundant wood supply and a good rail transportation network. The local data on labor rates and site conditions is available from the previous work and allows a reasonable comparison.

Potential changes occur in both plant capital cost and operational costs. The overall capital costs of the plant are virtually unchanged. Savings in overall construction labor costs and freight charges of approximately $3,000,000 are offset by the more severe climatic conditions. The changes are well within the accuracy of the initial estimate and no change in this area is anticipated. The changes in operating costs are summarized in Table 7.8-1. Lower costs for chemicals and electricity are partially offset by higher operating labor rates and lower electrical buy-back rates. The major operating advantage is seen to be in lower wood costs. The potential reduction in base case ethanol selling price is approximately $0.251/gal. This advantage could be easily offset by tax advantages of one location over another. It should be noted that $0.14/gal ethanol of this savings is due to lower wood costs which are already accounted for in the optimistic case in the sensitivity analysis.

TABLE 7.8-1

OPERATING COST COMPARISON OF
A CHANGE IN PLANT LOCATION[1]

Component	Unit Price Change	Cost Change $	Effect on Ethanol Selling Price (¢/gal)
Wood	-$9/ton	-2,085,714	-13.9
Sulfuric Acid	-$27.32/ton	-208,588	-1.4
Gasoline	-10¢/gal	-71,403	-0.5
Labor (base rate)	+$1/hr	+481,800	+3.2
Electricity	-.0278¢/kwh	-1,884,718	-12.5
		-3,768,623	-25.1

NOTE:

1. Hilo, Hawaii to Spokane, Washington

SECTION 8

TRADE-OFF STUDIES AND OPTIONS CONSIDERED

The base-case design was chosen according to the design criteria summarized in Section 4. It is stipulated that commercially available processing options and equipment be used where applicable. This criterion was established to provide a firm basis for plant capital and operating costs. The trade-off studies in this section examine some of the more promising alternative processing options to the base case which could lower the production costs of ethanol.

The trade-off studies and process sensitivities were carried out relative to the base case design. Rough material balances were calculated and the effect on energy integration, if any, was considered for each case, and equipment costs were obtained using appropriate scaling factors, based on equipment type. This type of analysis is valid within reasonable processing capacity or duty changes from the base case; however, the uncertainty increases at large differentials. The analysis discuss the various technical implications associated with each processing change; however, the economics for each trade off do not include any developmental or risk factors.

The results of the trade-off studies indicate that the recovery of valuable by-products from the C_5 fraction of the wood is the most significant way to reduce the production cost of ethanol. The production of furfural (assuming a selling price of $0.20 per pound furfural) can potentially reduct the ethanol selling price by approximately $0.45 per gallon ethanol. Alternatively, the production of ethanol via C_5 fermentation can potentially reduce the ethanol selling price by about $0.71 per gallon ethanol. Each of these options must be looked at in more detail before definitive conclusions can be drawn. Other methods to reduce the price of ethanol are also discussed in this section. Sensitivity studies on an increase in enzyme activity and improvements in hydrolysis efficiency indicate a potential reduction in the selling price of ethanol of by $0.18 and $0.21, respectively. These cases are only significant if the above mentioned by-product credit is obtained.

A process design incorporating the best trade-off alternatives was not prepared; however, a minimum ethanol price was developed by combining the additive trade-off differentials. These economics are discussed in Section 7. An optimal design case would require more definitive site-specific information, developmental data, and market research to determine which combination of alternative processing options would be most economic.

8.1 C_5 RECOVERY - BY-PRODUCT TRADE-OFFS

In the base-case design, the pentose sugars are recovered after steam explosion in a counter-current water wash. The extracted pentose sugars are sent to an anaerobic digester to produce a methane rich gas for boiler fuel. This processing sequence was chosen for the following reasons:

- Current markets for alternative uses of C_5 sugars (furfural production and animal feed) are not available in Hawaii.
- Inhibitory compounds must be removed before fermentation.
- Fermentation of the C_5 sugars to ethanol is not a proven or commercial technology.
- Burning the pentose sugars directly requires extensive capital and energy expenditures for dewatering and recovering volatile organics.

Evaluation of the base case economics indicates that a substantial credit from the C_5 function of the wood is required for commercial viability of the wood-to-ethanol process. Trade-off studies were performed to determine the relative merit of the several possible by-products.

The question of marketability of a by-product is one that has received a good deal of attention from proponents of the wood-to-ethanol process. The goal of these trade-off studies is not to quantify the markets for the particular C_5 product, but to determine the relative economics for C_5-product production if a market exists.

Studies by SERI (The Value of Furfural/Ethanol Co-production From Acid Hydrolysis Processes) have suggested that furfural is the most beneficial C_5 product to recover. The furfural can be produced from the C_5 wash stream by providing a separate furfural production reactor and recovery system prior to the anaerobic digesters. The reactor system can be operated to maximize furfural production and minimize by-product formation. Eighty percent of the theoretical yield of furfural from the C_5 stream is assumed for the trade-off study. Additional verification of furfural yields from extracted pentose sugar streams is necessary to quantify the results of this trade off. An additional advantage of producing furfural from the extracted sugar stream is the possibility of recovering acetic acid as a second by-product. In the base-case process design, the wash waters are recycled process effluents (evaporator condensates and beer still bottoms product); therefore, any acetic acid that is produced as a process by-product (i.e., from hydrolysis or fermentation) is accumulated in the wash stream. The acetic acid is not degraded in the furfural reactor and can be recovered as part of the furfural recovery system by using furfural as an extractive agent. Commercial processes are available for this recovery. This option is included as the primary furfural trade off for economic evaluation. Definitive testing is required to determine the quantity of acetic acid which could be recovered from the processed wood hydrolysate. Thus, the economic evaluation can be considered as an optimistic case.

The C_5 sugars can also be converted directly to alcohol via fermentation. C_5 fermentation is currently in the developmental stage;

therefore, the trade-off analysis is based on estimated capital cost and ethanol production values. It is assumed that inhibitory compounds, such as organic acids, furfural, and salts, will not affect the yields or fermentation times. Twenty-five hours of fermentation time with 75 percent of theoretical yield was assumed for the study economics.

The pentose sugars can be sold as an animal feed. The extracted stream is concentrated either by evaporation or spray drying, or a combination of both. This type of animal feed is commercially produced from defibrated wood extractives in the fiberboard industry.

Molasses waste is in abundance in Hawaii and would be in direct competition with the pentose sugars. Therefore, the selling price of the extracted pentose sugars would be $44/ton maximum. The economics for both the Hawaiian marketplace and an estimated continental U.S. market are included.

As with any feed commodity, the pentose sugars would require trace component classification, nutritional testing, and certification before a market could be established. Quantity lots of this by-product would be required for testing.

The economic incentives for those by-product options are discussed in Section 7.6.5, "Discussion of Results (Base Case)".

8.2 LIGNIN AS A BY-PRODUCT

Lignin is solubilized and removed from the main process steam in the base case design in a counter-current alkali wash. The recovered lignin is subsequently burned in a boiler to produce process steam. This processing option was chosen because an alternative market for lignin is not available in Hawaii and because of the technical impacts discussed in trade off in Section 8.7.

The economic evaluation of the base case wood-to-ethanol process indicates that additional revenue in the form of by-product credit(s) are necessary to attract private investors to this process. The goal of this trade-off study is not to define a market for lignin, but to determine the potential value of the lignin by-product, if such a market would exist. Table 8.2-1 shows the production rate and the values of lignosulfonates for the years 1978 to 1980. This table also shows that the value of the lignosulfonate by-product will vary with the processing method used for recovery.

It was assumed in the trade off that lignin was recovered in the same manner as in the base-case design. Additional processing requirements, if any, for sale of lignin were not included in the calculations. The selling price of lignin was assumed to be $0.15/lb net to the plant. For this analysis additional, supplemental wood was needed to replace the marketable lignin that was used as boiler fuel in the base-case design. This trade off resulted in a by-product credit of $0.69/gal ethanol produced.

In the base-case design, 83 million lb of lignin extracted with sodium is produced annually. According to the U.S. International Trade Commission (Table 8.2-1), the production of sodium-lignosulfonates in 1980 was approximately 102 million lb with total sales of 99 million lb, indicating a potentially saturated market. Although market analysis and surveys of lignin utilization processes show a growth potential in the lignosulfonate area, the potential for expansion of these markets and other lignin markets must be investigated in greater detail before any conclusion can be made as to the marketability of a lignin product.

TABLE 8.2-1

PRODUCTION AND VALUE OF LIGNOSULFONATES

	1978	1979	1980
Lignosulfonates, total			
Production (000s lb)	830,000	806,134	879,969
Sales (000s lb)	803,000	750,394	880,527
($ million)	52.000	55.780	63.761
Unit Value ($/lb)	0.06	0.07	0.07
Lignosulfonic Acids; Salts			
Calcium			
Production (000s lb)	554,000	590,131	620,890
Sales (000s lb)	526,000	540,524	625,386
($ million)	21.000	22.095	25.316
Unit Value ($/lb)	0.04	0.04	0.04
Sodium			
Production (000s lb)	121,000	99,765	101,983
Sales (000s lb)	121,000	97,742	99,356
($ million)	15.000	16.938	16.943
Unit Value ($/lb)	0.12	0.17	0.17
Iron			
Production (000s lb)	2,000	2,110	1,903
Sales (000s lb)	2,000	2,110	1,755
($ million)	0.368	0.368	0.321
Unit Value ($/lb)	0.17	0.17	0.18
Chromium			
Production (000s lb)	-	98,898	-
Sales (000s lb)	-	95,865	-
($ million)	-	15.326	-
Unit Value ($/lb)	-	0.16	-
Ammonium			
Production (000s lb)	-	13,941	-
Sales (000s lb)	-	12,864	-
($ million)	-	0.765	-
Unit Value ($/lb)	-	0.06	-
All Other			
Production (000s lb)	153,000	1,289	155,193
Sales (000s lb)	153,000	1,289	154,030
($ million)	16.000	0.288	21.181
Unit Value ($/lb)	0.11	0.22	0.14

Source: U.S. International Trade Commission, Synthetic Organic Chemicals, Annual Report

8.3 SODIUM HYDROXIDE RECOVERY

The lignin extraction section consumes 1443 lb/hr of sodium hydroxide, which is used to solubilize the lignin from the water washed steam exploded wood. The feed alkali is about a 1.2 wt percent caustic solution. In the base-case design, the solubilized lignin is recovered by pH adjustment with sulfuric acid. At about pH 9.0, lignin becomes insoluble and can be separated by conventional means. The disadvantage of this system is that large quantities of both caustic soda and sulfuric acid are consumed and the liquid waste stream will contain sodium sulfate which creates a disposal problem.

An alternative method for reducing the chemical and disposal costs is represented in process flow sketch, Figure 8.3-1.

The improvements are:

1. Neutralize with a stronger alkali solution (about 6 wt percent). This will permit more washwater to be used in the upstream wash system, thus reducing the number of wash stages.
2. Recover the spent alkali by reducing the pH to about 9 with CO_2 available from the fermentation section and adding lime to regenerate sodium hydroxide for recycle to the wash section. Thus:

$$NaOH + \text{lignin} \rightarrow Na\ (\text{Lignate}) \quad (1)$$

$$Na\ (\text{Lignate}) + CO_2 + H_2O \rightarrow Na_2CO_3 + NaHCO_3 + \text{soluble lignin} \quad (2)$$

$$2\ NaHCO_3 + Ca(OH)_2 \rightarrow CaCO_3 + Na_2CO_3 + 2H_2O \quad (3)$$

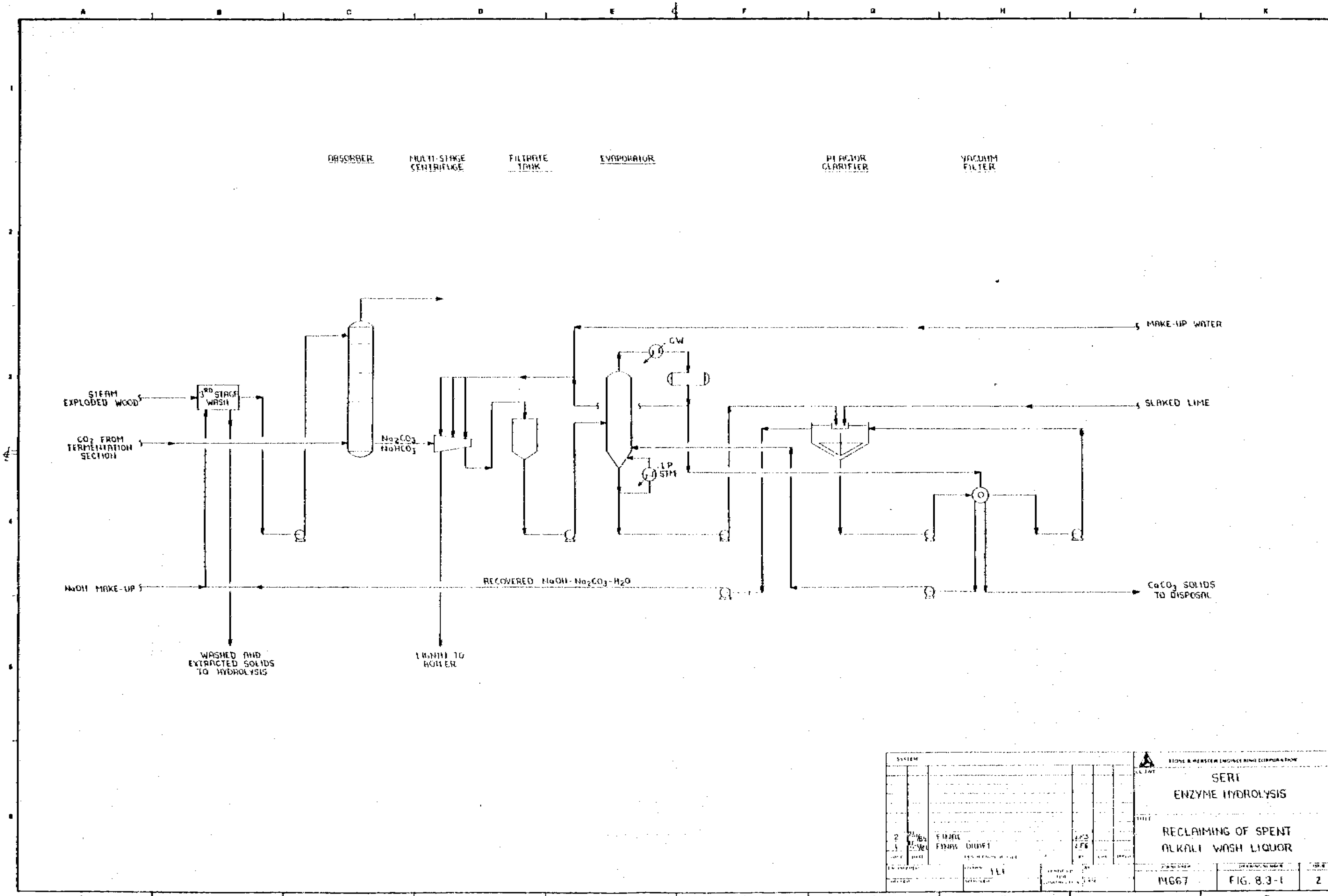
$$Na_2CO_3 + Ca(OH)_2 \rightarrow CaCO_3 + 2\ NaOH \quad (4)$$

Lignin is solubilized in the caustic wash with a $NaOH$-Na_2CO_3-H_2O solution. The resultant liquor is then acidified to a pH of about 9 with waste CO_2 from the fermentation section. At pH of 9, the CO_2 forms a mixture of carbonate-bicarbonate while precipitating the lignin. The lignin is separated from the carbonate-bicarbonate liquor via centrifugation. Alkali losses are reduced by thorough washing of the lignin. The lignin is then used as boiler fuel, or may be sold for other uses if it is of a reactive nature.

The filtrate goes to a water removal step which is required so that the system water balance is maintained. A water imbalance is caused by the use of wash water, for alkali recovery in the solids separation stages, which contains the recovered alkali. Continued recycle of wash water without purification would result in alkali build-up in the wash liquor which would preclude its use. After concentration, the liquor is reacted in a reactor-clarifier with slaked lime to form insoluble calcium carbonate. The calcium carbonate is separated from the causticized mother liquor in a vacuum filter and then washed or reslurried to mimimize the alkali loss. The regenerated caustic solution is recycled to the lignin alkali wash stage.

Reaction of lime with sodium carbonate has been practiced for years in the pulp and paper industry, and in the heavy alkali chemical industry. The major difficulties in this system involve minimizing the alkali loss in the solid separation step and obtaining high reaction rates and conversion efficiencies.

Extensive R/D work is not required for the caustic regeneration system. Centrifuge wash rates, settling rates, and reaction conditions have all been commercially demonstrated. This information is available from equipment, vendors and literature sources. However, additional demonstration would be required to verify the design of the lignin precipitation recovery equipment, as well as determine the solubility of lignin in highly carbonated solutions. The cost advantage for this system is about $0.15/gallon of ethanol. This system should be included in any future conceptual design of this plant concept.



ABSORBER
MULTI-STAGE CENTRIFUGE
FILTRATE TANK
EVAPORATOR
REACTOR CLARIFIER
VACUUM FILTER
STEAM EXPLODED WOOD
3RD STAGE WASH
CO2 FROM FERMENTATION SECTION
Na_2CO_3 $NaHCO_3$
CW
LP STM
MAKE-UP WATER
SLAKED LIME
NaOH MAKE-UP
RECOVERED NaOH-Na_2CO_3-H_2O
$CaCO_3$ SOLIDS TO DISPOSAL
WASHED AND EXTRACTED SOLIDS TO HYDROLYSIS
LIGNIN TO BOILER
SYSTEM
FINAL
FINAL DRAFT
SERI
ENZYME HYDROLYSIS
RECLAIMING OF SPENT ALKALI WASH LIQUOR
M667
FIG. 8.3-1
2

8.4 EFFECT OF HYDROLYSIS YIELD, RESIDENCE TIME, AND CONCENTRATION

The design criteria of the hydrolysis of cellulose to glucose was based on work done at Berkeley. Data from various sources indicate that, at cellulose concentrations above 5 wt percent, the hydrolysis residence time would be approximately 48 hours. This reaction time is necessary because of glucose inhibition of the enzyme complex and the buildup of the intermediate products, cellobiose and other reducing sugars.

The base case design assumed a cellulose concentration of 7 wt percent producing an 84 percent yield of fermentable sugars in 48 hours. This trade-off study considered an increase in the cellulose concentration to 10 wt percent, sugar yields of 90 percent, and a residence time of 24 hours to assess the economic incentives for pursuing R&D efforts to achieve these design goals.

Due to the increase in fermentable sugar yield, the trade off shows a 6.4 percent decrease in the wood feed to the process. This wood reduction diminishes the capital cost of the pretreatment and enzyme production areas (sections 100-300) by about 4 percent. The hydrolysis section's capital cost was decreased by 50 percent. This reflects the combined effect of the increased yield, lower residence time, and the higher cellulose concentration. The increased cellulose concentration reduces the evaporator capital and utility costs by approximately 19 percent and 49 percent, respectively. The decrease in the evaporator steam consumption reduces the LP steam requirement. This LP steam reduction, coupled with the decreased HP steam requirement for the steam explosion guns, reduces the boiler capital cost by 13 percent.

The reduced feedstock requirement lessens the amount of methane and lignin going to the boiler as fuel; however, supplemental wood to the boiler is not increased because of an equivalent decrease in steam consumption.

The trade off shows that if a 10-percent cellulose concentration, a 90-percent yield of fermentable sugars, and a hydrolysis residence time of 24 hours is achievable, a reduction in ethanol selling price of $0.21/gal could be reached.

8.5 EFFECT OF INCREASE IN ENZYME ACTIVITY OR DECREASE IN ENZYME FERMENTATION RESIDENCE TIME

The base-case design assumes that enzyme requirements are met by in-plant production utilizing the Rut-C-30 Trichoderma organism. Thirteen days of residence time are provided to produce 30 filter paper units from a 15 wt percent cellulose broth. Using this criterion, base-case capital and operating expenses for enzyme production account for $0.48 per gallon of ethanol or 13.8 percent of the required ethanol selling price. This cost consists of approximately $0.12 in capital-related charges, with the remaining $0.36 directly attributable to operational expenses (not including labor and common facilities expenses). This cost compares to enzyme charges of about $0.05 per gallon in grain ethanol plants.

There are various methods to reduce enzyme production costs. These include increase of enzyme activity, increase of enzyme titre, reduction of fermentation time, higher enzyme recovery rates (i.e., enzyme stability), enzyme immobilization (enzyme viability), and enzyme production using a less costly feedstock. Of these, increasing the enzyme activity and reducing the fermentation time provide the best near-term alternatives and were evaluated as trade-off studies. The first trade off evaluates the cost effect of a two-fold increase in enzyme activity, and the second a 50 percent reduction in fermentation (residence) time. In both studies, the major cost impact occurs from a reduction of air sparge and nutrient requirments.

The capital and utility cost reduction in the enzyme production section (300) is 50 percent of the base-case enzyme production capital cost. The decrease in residence time will also reduce the fermentation refrigeration load and result in a capital and utility savings of approximately 3 percent and 22 percent, respectively. The increased enzyme activity case has the added incentive of a reduction in process feedstock. This reduction in wood feedstock decreases both the capital and utility costs of the pretreatment sections (100 and 200) by 2.4 percent. A decrease in the process feedstock will decrease the production of methane and the quantity of lignin used as fuel in the boiler. In order to satisfy the process steam consumption, supplemental wood to the boiler is required. The net change in total wood feed is a 1.2 percent reduction.

The economic incentives for doubling enzyme activity and halving enzyme fermentation time are $0.18 and $0.09 per gallon ethanol, respectively.

8.6 EVAPORATION - EFFECT ON COSTS

The evaporation system is designed to concentrate the hydrolyzed eucalyptus wood sugars from 5.7 wt percent to 14.7 wt percent fermentable sugars. In addition to providing a concentrated sugar stream which can be processed in commercially available fermentation and distillation systems, the evaporation system also serves to remove acetic acid, furfural, and other volatile organic fermentation inhibitors from the fermentation feed. The recovered evaporator condensates provide a low salt process water which can be reused as internal plant recycle. These processing advantages are offset by the higher capital cost and steam requirements for evaporation.

This trade-off study evaluates the economics of removing the evaporation system from the process. Neither the inhibitory effects on the fermentation due to recycled impurities nor the accumulation of volatile organics in the distillation system were considered in this evaluation. Testing at a pilot level would be necessary to demonstrate the operability of the fermentation and distillation units when processing wood hydrolysate streams.

Elimination of the evaporators affects the operation of downstream equipment. The impact on the process economics requires an evaluation of the following:

1. Fermenter cost, operation, and yield when fed a more dilute sugar stream.

2. Beer still size, energy requirements, and operation on low alcohol concentration feeds.

3. The effect of substituting "dirty" recycle water (beer still bottom) for the recycled evaporator condensates.

The capital cost of the installed evaporation system, including control room and other charges, is approximately $2.2 million. This amount was eliminated from the base-case costs. An additional cost of $75,000 is incurred in the distillation system which involves modifications to the beer still and heat exchangers. The beer still diameter increases, but the height remains the same. The reflux ratio was increased to maintain the same number of rectifying and stripping trays in the column. An evaluation of closer approaches to minimum reflux (lower energy requirements) at the expense of more tower trays was performed and found to have marginal benefit. The beer still overhead condenser and preheater sizes increase and a beer still bottom cooler is added to cool the recycle water.

Because of the increased liquid loading at the higher reflux ratio, the beer still required an additional 35,000 lb/hr of low pressure steam. This additional steam when offset by elimination of evaporator steam gives a net boiler steam reduction of 24,000 lb/hr. The base case LP steam demand exactly matches the steam required for the enzyme air compressor (R-301A,B) turbine drives. With a lower LP steam demand, this would no longer be true. One way of maintaining steam drives for both air compressors at the lower LP steam demand is to install a condensing turbine for one unit and an extraction turbine to provide the 50 psia LP steam demand. For this case, it is estimated that the boiler wood (fuel) requirement would drop by about 4,500 lb/hr.

The effect of a more dilute feed on the immobilized bed fermentation system is difficult to quantify without the benefit of pilot-scale data on wood hydrolysates. However, it was assumed that the result of greater liquid flow rates and reduced residence time because of lower sugar concentrations and ethanol inhibition would yield a constant fermenter space velocity (cubic feet of fermentation volume per ethanol production). Therefore, no capital adjustment was included for change in the fermentation system. This is obviously the most optimistic case.

The key variables to consider in ethanol production and yeast cell maintenance and growth are:

- Type and concentration of feedstock
- Concentration of salts
- Ethanol concentration
- By-product concentrations
- Temperature and pH
- Inhibitory components

For a continuous fermentation system, the reactor space velocity that produces the maximum ethanol productivity is determined by the optimum operating point. Since ethanol, CO_2, and sugar concentrations will be different at 15 wt percent glucose feed than at 5 wt percent glucose feed, the reactor space velocity probably will be different at the lower concentration for the maximum ethanol productivity. Reactor mechanical design considerations should also be considered. Kyowa notes in its general literature that "it was recognized that at least two columns must be connected in series to obtain higher conversion owing to the strong turbulent effect of carbon dioxide evolved during fermentation." In short, the effects on fermentation capital costs in the more dilute region must be determined via experimentation or from vendor experience.

A most important benefit of the evaporation system is that it provides a clean condensate which can replace clean sterile process water makeup. Decrease in makeup water reduces total plant effluents and, therefore, treating costs. This is a major capital item in the plant costs. An increase in treatment cost or well drilling expense was not included in this tradeoff. In the trade off design, it has been assumed that beer still bottoms can be recycled as process makeup water. The beer still bottoms contain many impurities and the effects of these recycled impurities (primarily high salt loadings in enzymatic hydrolysis) could be deleterious to the process. Process effects caused by recycling beer still bottoms must be established.

The plant economics show about a $0.06/gallon ethanol advantage when eliminating the evaporators. It should also be recognized that potential cost increases in environmental treatment and water makeup, if the beer still bottoms can not be recycled, may eliminate this savings. The savings must be weighed against the qualitative operating advantages of installing evaporation equipment.

8.7 LIGNIN REMOVAL

In the base-case design, lignin is extracted from the cellulose/lignin complex in an alkali wash system before enzyme production or cellulose hydrolysis (Section 5.1). This option was chosen for the base-case design for the following reasons:

- The presence of lignin during hydrolysis may have an adverse effect on the rate of hydrolysis and overall glucose yield. The lignin still associated with the cellulose is expected to shield or block a portion of the cellulose from enzymatic attack.

- The higher solids concentration due to the presence of insoluble lignin in both the enzyme fermenter vessels and the hydrolysis reactors may impose a constraint on vessel agitation design and cause higher equipment capital and operating costs.

- The presence of lignin is expected to interfere with enzyme recovery and recycle. It has been noted that lignin may act as an absorbent for enzyme. A portion of the cellobiohydrolase and endoglucanase (C_1C_X) enzyme would therefore be absorbed onto the lignin solids present during hydrolysis and be removed with the centrifuged cake.

- An extracted lignin is considered to be more reactive and, therefore, a more valuable product.

The technical and operating incentives for lignin removal before hydrolysis are counterbalanced by the high capital and operating costs associated with the lignin extraction equipment.

This trade off evaluates the economics of removing the lignin after enzymatic hydrolysis rather than removing the lignin prior to hydrolysis as in the base case. Neither the impact on the rate of hydrolysis and glucose yield nor the effect on agitation design was considered in the evaluation. It is necessary, however, that the impact of lignin presence during cellulose hydrolysis and enzyme production be demonstrated through pilot plant testing.

Removal of lignin after hydrolysis will have an effect on enzyme recovery. It was found that enzyme adsorbed on the lignin was lost with lignin in the hydrolysis centrifuge cake. This loss results in a need for 34-percent increase in the enzyme production requirements. The increase in enzyme production increases the wood feed to the process by approximately 2-1/2 percent and increases the quantity of nutrient required for enzyme production.

The major economic effects of recovering lignin after hydrolysis and enzyme production are:

- An increase in installed capital cost of the enzyme production unit from $6.2 million in the base case to $7.6 million.

- A savings of $1.6 million by removal of the alkali wash system.

- An increase in the anaerobic digestion installed capital cost of $0.43 million.

- A decrease in capital cost of the lignin centrifuge of $0.43 million because of a reduction in lignin centrifuge capacity caused by the elimination of the caustic/wash water required in the counter-current alkali wash system.

It has been assumed that 95 percent of the soluble solids can be recovered in the 5-stage counter-current water wash (V-401) following the hydrolysis centrifuge. Pilot plant testing of the counter-current water wash system is necessary to demonstrate the ability to recover this high percentage of soluble solids from the hydrolized lignin cake. The inability to achieve this 95 percent recovery level will increase sugar losses and water rates, resulting in a significant cost impact on the trade-off study. Removal of lignin after hydrolysis did increase the wood feed to the process, however, the net wood to the plant, process feed plus supplemental wood to boiler, did not change significantly from that of the base case. This is because anaerobic digester methane production increased to satisfy net steam requirements.

A savings of about $0.21/gallon of ethanol produced can be achieved by removal of lignin after hydrolysis. This savings is a result of the elimination of sodium hydroxide used to solubilize lignin in the alkali wash prior to hydrolysis. Additional pilot demonstration is required to determine the operability of the process with lignin present in the hydrolysis and enzyme production sections. Consideration of the fact that recovery of the sodium hydroxide, used in removal of lignin prior to hydrolysis, has the potential savings of about $0.15/gallon (see trade off, Section 8.3) indicates that additional research for this method should probably not be pursued at this time.

8.8 PRETREATMENT REQUIREMENTS

The base case design assumes that approximately 48 hours of residence time for acid impregnation is required to enhance the yield effects of steam explosion at moderate steam explosion conditions. This residence time is based on relative data obtained by Iotech (1982) while under contract with the Department of Energy.

This trade-off study evaluates the economic effect of changing the method of presoak. The presoak alternatives include: pressure soaking, reduced soak residence time, and presoak elimination by acid spraying prior to steam explosion.

The first option for evaluation is the combined effect of pressure presoak and reduced residence time. The trade off assumes that soak time can be reduced by one-half and that additional capital costs for pressurization are not required.

Reducing the cycle time from 48 to 24 hours does not mean that actual time that the chips are in contact with sulfuric acid is reduced by one-half. The basis of design assumes a 48-hour cycle time, of which about 34 hours is actual chip-acid contact time. The remaining time is consumed for bin filling and unloading. There are many schedules that can be developed for a given timed cycle. The required task is to develop a reasonable cycle that results in reasonable equipment sizes. In developing a reasonable 24-hour cycle for the chip impregnation step, the most obvious manner to accomplish this was to reduce the number of bins from 6 to 3, or to process the material twice as quickly. Since the chip filling and liquid loading and unloading times are the same for both the 48- and 24-hour cycle, then the actual soak time would be reduced from 34 hours to 10 hours, which is about a two-thirds reduction. The cost effect of this cycle reduction time lowered the Section 100 capital cost from $2,500,000 to about $1,350,000.

Another scenario for a 24-hour impregnation cycle was to reduce the number of bins from 6 to 4 and their size from 43,000 ft^3 to 33,000 ft^3. In order to maintain a constant flow of material into and out of this section without intermediate storage, it is necessary that the chip fill time equal the chip discharge time. Fixing this time at 6 hours results in about 18 hours remaining for chip impregnation with acid. The cost effect of this cycle reduction time also lowered the Section 100 capital cost from $2,500,000 to about $1,350,000.

Thus for about the same reduction in capital cost achieved by two different designs, it would be possible to obtain an 80-percent higher chip impregnation time (10 hours versus 18 hours). In the latter case, the chip impregnation time was reduced to about 50 percent of the base-case design (34 hours versus 18 hours). The net economic effect of chip soak residence time reduction is $0.02/gal ethanol.

A second option for consideration is the elimination of acid presoaking section. The presoak section is replaced with a system for spraying acid on the wood chips immediately before steam explosion. The trade off assumes that capital costs in the steam explosion section will not increase as a result of the acid spraying equipment. A reduction of $0.04/gallon is realized in this case.

8.9 MOLECULAR SIEVE DEHYDRATION OF ETHANOL

The base-case ethanol dehydration process uses azeotropic distillation to convert the nominal 190-proof ethanol to anhydrous quality. Molecular sieve technology is also available for ethanol deyhydration. This economic evaluation of vapor phase ethanol dyhydration using molecular sieves shows that the molecular sieve process reduces the investment and operation costs for the dehydration section, compared to azeotropic distillation. An advantage of $0.03/gallon of ethanol is possible. The advantages of using a molecular sieve process compared to the azeotropic distillation include:

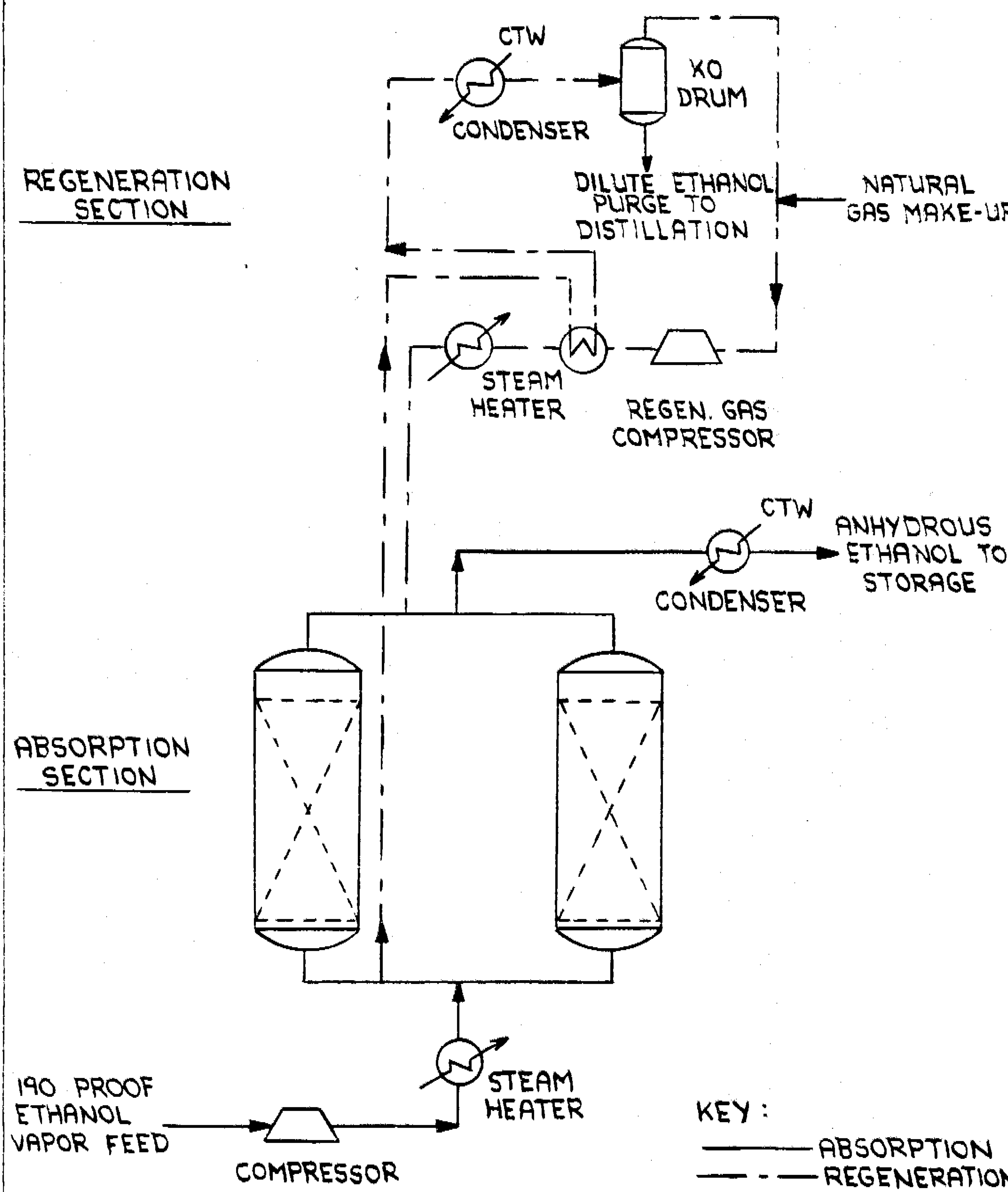
- The process offers better stability and is more resistant to upsets
- Simple design, minimum structures and foundations
- Sieve performance is guaranteed by sieve vendor to eliminate performance risks

The vapor phase molecular sieve dehydration simplified process flow diagram is shown in Figure 8.9-1. The system consists of an absorption section and a regeneration section. The absorption section is where the 190-proof vapors from the distillation column are compressed and sent to the molecular sieve dryers for water removal and anhydrous ethanol production. The anhydrous ethanol vapors are then condensed and stored in the product tanks. After sieve bed water breakthrough has occurred, the regeneration section is used to regenerate the dryer. This section uses a closed-loop system to heat the regeneration gas in order to strip the water from the molecular sieve, then cool the regeneration gas to condense and remove the water-ethanol purge for refractionation and ethanol recovery. This recycle stream is about 2 percent of the initial fractionation capacity.

The actual temperatures, pressures, and flow rates used in the absorption and regeneration cycles are proprietary to the sieve vendor at this time. The public information available for disclosure is shown in Figure 8.9-1. The vapor phase dehydration is economically more attractive than the liquid phase molecular sieve process because the quality of sieve required is lower, the dryer vessels are smaller, the temperature variations on the sieves are lower, the operating costs are lower, and the investment is lower.

The economic incentive for vapor phase molecular sieve dehydration amounts to a saving of about $0.03/gallon ethanol over the base-case azeotropic distillation system. The inclusion of molecular sieve dehydration would require a more detailed evaluation of site-specific economics, operating advantages, and demonstrated commercial experience.

MOLECULAR SIEVE DEHYDRATION OF ETHANOL



NO.	REVISION DESCRIPTION	BY	DATE		
2	FINAL	LFS	2/1/85	SCALE / DESIGN	STONE & WEBSTER ENGINEERING CORPORATION
1	FINAL DRAFT	LFS	9/13/84	CHECKED	SERI - ENZYME HYDROLYSIS
CERT. FOR FAB.	CERT. FOR CONST.	APPROVED		DRAWN LEF	AREA: / J.O. NO.: 14667 / DRAWING NO.: FIG. 8.9-1 / ISSUE: 2
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8.10 WOOD HANDLING - STACKER VS DUMPED PILE

Two methods of wood handling were considered for chip storage and reclaiming for this plant. The design case is an automated overhead stacker/reclaimer system as described in Section 5.1. This system will ensure a first-in/first-out (FIFO) chip turn-over and 100 percent rotation of the piles.

An alternative wood handling option, dumped pile, consists of a belt pile builder, bulldozers, and a drag chain or screw reclaim conveyor. Chips are conveyed by belt from the receiving station and distributed to a single pile with a rotating mechanical thrower. Bulldozers operating 24 hours per day continuously compact and move chips to the reclaiming area.

The automated FIFO chip management system used in the stacker option can result in significant wood savings over the dumped pile option considered, which cannot ensure FIFO. Automatic chip management systems with FIFO are especially effective in reducing pile shrinkage (i.e., wood losses) for large wood users (above 10 tph). Wood losses or pile shrinkage in open wood chip storage piles are a result of anaerobic bacterial degradation with the major portion of these losses occurring during the first 90 days of storage. Hence, a reliable first-in/first-out system with 100 percent rotation of piles is desirable. Studies have shown that an overhead stacker/reclaimer system can be economically justified based primarily on wood savings for 800 tpd pulp mills (approximately 90 wet ton/hr feed rate) (Chieves 1977). In addition, the FIFO system will minimize the potential for wood pile fires resulting from poor turnover. The major economic savings using the stacker/reclaimer are:

- Lower manpower requirements
- Lower energy requirements

The manpower and energy requirements associated with the stacker option for this plant size are expected to be significantly lower than those associated with the dumped pile system. A single operator can control the operation of the stacker system from a remote tower, whereas the dumped pile option will require two dozer operators and a pile building/reclaim operator for each shift.

The energy consumption of the stacker system is typically lower than that for the dumped pile due to the high fuel requirements of the bulldozers. Schleger and Jepsen (1978) show this lower energy requirement in a comparison of the energy costs for stacker vs dumped piles with a pneumatic feed system. There should be significant energy savings associated with the stacker option considered for this size plant.

Other advantages of the stacker option over the dumped pile option are:

- Increased reliability - longer downtime is typically associated with drag chain conveyor maintenance, pile building, and dozer maintenance.

- Potential to blend the chip pile, if necessary.

- Minimal chip damage.

- Less chip contamination from rocks, dirt, and dozer rubble.

The stacker method of wood handling is recommended for this plant. Wood savings alone can justify the investment in this system over a dumped pile system. Other advantages include lower manpower and energy requirements and increased reliability.

8.11 WOOD SIZE REDUCTION

It is necessary to reduce the size of the nominal 3/4-inch wood feed chips to make the cellulose in the wood more accessible to enzymes for hydrolysis. Before the wood can be enzymatically hydrolysed to fermentable sugars, there must be direct physical contact between the cellulose fibers and the enzyme. In addition, it has been found that some of the compounds found in wood and its hydroloysates can have inhibitory effects on either the enzymatic hydrolysis of cellulose or the fermentation of the derived sugars. Therefore, it is important that the size reduction/pretreatment process both render the cellulose accessible and assist in making the other inhibitory by-products extractable from the cellulose.

Several types of size reduction/pretreatment processes were considered for this trade-off study, including grinding, milling, defibration, and steam explosion.

Stone grinding is a technically viable option for size reduction. However, the pulp and paper industry has found that it is energy- (requiring 1300-2800 kWh/BDT) and labor-intensive. Inclusion of pretreatment equipment would also be required to remove by-products that can inhibit enzyme production, hydrolysis, or yeast fermentation. In addition, this option would require whole log type feed and completely different wood handling storage and delivery systems. This option was not given further consideration for the reasons stated above.

Milling of wood chips is accomplished commercially using hammer mills. However, these mills can only reduce the size of the wood to 30-mesh, due to strength limitations of the milling screen. The mesh size is larger than the size material used to generate the design criteria data. Testing of the larger particle size wood in enzyme production and hydrolysis would be necessary. This option also requires that the wood chips be dried to 9-percent moisture to lower the milling electrical power requirement (50 kWh/BDT). A preliminary estimate of the cost of milling has been made, assuming flue gas can be used for drying the wood. The energy and capital costs were found to be about $0.25/gallon ethanol higher than those developed for steam explosion. Due to the increased costs and the inability of this process to meet the size and quality obtainable by other processes, this option was not selected.

Defibration is a commercially proven method of size reduction for wood chips, in which wood fibers are separated in a disc refiner. The defibration is typically accompanied by steam pretreatment, and is included in this trade-off study because it significantly reduces the electrical power required in the refiner (43 kWh/BDT) and also makes the lignin and hemicellulose more extractable.

In the process of steaming, an autohydrolysis of the sugars takes place, predominately of the hemicellulose. Wayman (1978) has reported values in excess of 80 percent removal of the hemicellulose from wood in a sequential wash step using this process. He also found that the resulting washed pulp is suitable for enzyme hydrolysis. Similar processes have been used commercially in the pulp and paper industry since the 1930s and are currently being used in the manufacture of thermomechanical pulp (TMP).

This process with steam pretreatment requires much less energy than does grinding or milling. Approximately 50 percent of the steam required in the process, plus that generated by the refiner, can be recovered. Preliminary estimates show a $0.07/gal ethanol cost increase for this option over the steam explosion method.

The steam explosion process uses steam to heat the wood under high pressures to the point of structural softening. Then the pressure is released quickly, exploding the wood, rendering the cellulose fibers more accessible to hydrolysis. Steam explosion is being used commercially for the production of masonite board. Several similiar steam explosion processes have been developed for pretreating woody biomass before hydrolysis; the best known is the IOTECH process. The steam explosion process has been studied extensively to develop the optimum steam explosion conditions (time, temperature, pressure) for enzyme hydrolysis, although no commercial-scale plants are currently in operation.

This process offers several technical advantages over other size reduction processes. Experimental hydrolysis data clearly indicate that cellulose is made more accessible to enzyme hydrolysis than untreated ground wood. The lignin is rendered soluble in a dilute alkali wash for recovery. A high percentage of the hemicellulose is rendered water soluble. The energy requirements are lower than that required for milling and grinding, and about the same as the defibration option.

Steam explosion was selected as the method best suited for size reduction prior to enzymatic hydrolysis. This option offers the most favorable economics and can result in the highest hydrolysis and fermentation yields by rendering the cellulose accessible and the other fractions of the wood extractable.

SECTION 9

ASSESSMENT OF ENZYME HYDROLYSIS PROCESS

9.1 INTRODUCTION

This study evaluated the economic feasibility of producing ethanol via the enzyme hydrolysis of wood for the site specific case of Hawaii. SWEC based its integrated process design around unit operations that are in the bench-scale stage of development (e.g., enzymatic hydrolysis and enzyme production), and pilot-scale stage of development (e.g., steam explosion). The remaining process operations used commercial type equipment in which the precise operability (if unknown) was assumed to be optimistic, even though the exact composition of some of the process streams was not known to the degree required for a design construction. The major technical and mechanical uncertainties, discussed in Section 9.1, must be resolved by testing to confirm the base case design.

The economics of this process require significant revenue from the pentose (C_5) stream (either furfural or ethanol) and lignin stream to gain investor support. The initial objective of any future work is to quantify the potential markets and prices for these by-products. This step can determine if continued effort on the other technical parameters is justified. The R&D consists of both market definition and potential technical advances. A technical R&D program should be built on a firm economic basis.

9.2 TECHNICAL AND MECHANICAL UNCERTAINTIES

The base-case cost estimate was derived from equipment and vendor design specifications. These specifications were based on stream properties, flow characteristics, mixing properties, and separation criteria, assuming that the steam exploded wood and hydrolysate slurries would have characteristics similar to those found in commercial pulp and paper, food processing, or grain-based ethanol facilities. Even these properties are often determined empirically. More specific information on thermal conductivity, solubility, viscosity, and other fundamental properties of slurries, as well as operational data, is necessary to formulate a detailed design for construction or a definitive cost estimate.

Solids handling is an area where additional information is required to firm up the engineering design basis. The enzyme hydrolysis plant includes numerous belt and screw conveyors and high solid/liquid ratio pumps. In addition, there are areas where gravity discharge of high solids solutions from tanks, flash vessels, and transfer chutes is assumed. Furthermore, many of these streams could require sterile or septic service. Better characterization of these wood and hydrolysate streams by collection and correlation of pilot- and bench-scale data, as well as determination of sterility requirements, is required. The pentose sugar recovery, lignin wash, and hydrolysis enzyme recovery units utilize solids separations equipment (centrifuges, belt filters, cyclone, etc). The separation parameters assumed in the base case affect the recovery efficiency, wash rates, and ultimately downstream processing equipment. These separation parameters must be verified on vendor equipment so that definitive design parameters may be obtained.

The plant materials of construction have been selected, based on existing data. The choice of materials of construction is often dependent on the trace components found in the process streams (i.e., sulfur compounds, chlorides, oxygen levels, carbon dioxide, organic acids, etc). The effective selection of materials would require more accurate accounting of these species along with appropriate temperature criteria.

The hydrolysis and enzyme production sections of the plant contain large stirred reactor vessels, with the enzyme vessels being air sparged. The data used for the design of these vessels and the mixing requirements were scaled from laboratory data. The scale-up resulted in significantly lower mixing power and air sparge rates on an equivalent reactor volume basis than would be applied in bench scale apparatus. Pilot- or bench-scale testing at these operating conditions is required to determine the effect of process scale-up.

9.3 PROCESS UNCERTAINTIES

The major process uncertainty in the enzyme hydrolysis process is the applicability of eucalyptus wood to enzyme production, enzyme hydrolysis, and subsequent fermentation of the glucose to ethanol. The process has been designed to reduce, as much as possible, the effect of the inhibitory compounds found in eucalyptus. This is the primary reason for the inclusion of the water and caustic washes (lignin removal) and evaporator systems. Conclusive testing of eucalyptus feedstocks is necessary to determine the design plant yields and effects of inhibitory components. The base-case design assumes a specific eucalyptus wood composition (see basis of design, Section 4). This composition is typical of Eucalyptus globulus. As with any wood feedstock, variations in chemical composition between species and tree stands will appear. Selective sampling and analytical testing to determine feedstock variability and its effect on plant design criteria and operation are required.

The anaerobic digestion and the waste treatment sections greatly influence the overall plant economics. An increase in the conversion efficiency in the anaerobic digester would lower the production cost of ethanol. An in-depth study of plant effluents from a pilot unit would help quantify the effluent treating costs and effect on plant operation.

9.4 AREAS REQUIRING TESTING

As mentioned in the Introduction (Section 9.1), the enzyme hydrolysis process is in the bench/pilot scale stage of development. Areas requiring testing can be divided into three sections: areas where vendor equipment requires demonstration for final design and vendor guarantee; areas where integrated pilot plant operability and reliability are necessary; and areas which require verification of laboratory data and scale-up parameters. The latter two items should be obtained from a pilot-scale unit which would then have a dual function of:

1. Providing material for vendor testing

2. Providing data for commercial plant design

The following is a list of vendor testing requirements:

Equipment	Test
1. Impregnation Discharge System	- Determine a viable method of soaked chip discharge that will prevent solids bridging
2. Water/Alkali Wash System	- Determine the extraction efficiency and wash rates required for the recovery of pentose sugars and lignin
3. Centrifuge	- Determine the efficiency and operability of solid bowl centrifuges using hydrolysate slurries and extracted lignin streams.
4. Stirred Reactor Vessels	- Determine the mixing and air sparging requirements for enzyme production and hydrolysis reactors with and without lignin extraction
5. Hydrolysis Counter-Current Water Wash	- Determine the recovery of soluble solids from a hydrolyzed stream containing lignin.
6. Fermentation	- Determine the fermentability of the hydrolysate (rates and efficiency) at varying sugar concentrations and the effect of recycled water impurities
7. Anaerobic Digestion	- Determine the digestibility of the pentose/waste streams, including rates and efficiency at varying solids loading.

The following is a list of data necessary for future work:

Data	Description
1. Feedstock Characteristics	- Determine the componsition, variability, and processing properties of eucalyptus wood.
2. Slurry Properties	- Determine the flow characteristics and physical properties (e.g., viscosity, thermal conductivity, heat capacity) of a range of slurry concentrations.
3. Corrosion and Erosion Data	- Determine the corrosivity and erosivity of the pretreated wood, acidic slurries, and hydrostate streams. This data will be used to determine both contamination and material selection.
4. Scale-up Data	- Determine the important scale-up parameters and collect data to aid in scale-up of the hydrolysis and enzyme production sections.
5. Inhibiting Species and Levels	- Determine the level of toxicity of the impurities to types of fermentation yeast and waste treatment microbes. Determine the effects of recycled water impurities on enzyme production and hydrolysis.
6. Separation Parameters	- Determine the specific gravities and setting velocities for various concentrations of hydrolysate and steam-exploded wood slurry.
7. Neutralization Requirements	- Determine the level of natural base and organic acid in the hydrolysate and the amount and residence time necessary for sufficient neutralization.

9.5 RECOMMENDATIONS FOR FUTURE RESEARCH AND DEVELOPMENT

The enzyme hydrolysis process design is in the bench and pilot-scale stage of development. The economics of this study show that significant revenue from the pentose and lignin by-products is necessary to make this process economically viable. Future research and development should be directed to the markets and values of both pentose and lignin. A suitable feedstock that will minimize cost and inhibitory compounds and maximize total revenue should be defined, if possible. The justification of future technical improvements depends on the verification of a firm economic basis.

The high degree of interaction among the various process improvement options makes it difficult to prioritize research and development needs. The viability of these options is derived from a combination of economic and technical assumptions (see trade offs, Section 8). The enzyme hydrolysis research and development needs are given in a relative order of importance in Table 9.5-1.

Feedstock Research

The suitability of any feedstock for the enzymatic hydrolysis plant is dependent on cost, availability, and the quantity of potential by-products that must be sold. The portion of the overall operational costs associated with eucalyptus costs is $0.566/gal. Substitution of alternative feedstocks must not only be looked at considering potentially lower costs, but also by the amount of pentose and lignin by-products that must be sold to achieve positive economic results. In addition, the amount of total feedstock available must be compared with the required production of products to justify the research and development effort that must be expended.

The economics of wood as a feedstock for the production of ethanol should be compared to the economics of other potential lignocellulose feedstocks, such as bagasse, MSW, pulp and paper wastes, corn stover, etc. The composition of various feedstocks should be determined to identify existing inhibitory compounds and impact on waste disposal. The price of the feedstock will reflect its availability and market demand, as well as its harvesting method and handling charges. An economically attractive feedstock may be one that is currently considered as a waste stream from an existing process. An example would be the fines or particulate sludge from a pulp and paper mill. This waste stream has been collected and delignified within the paper mill process. Utilization of this material as a feedstock has the potential to eliminate the pretreatment sections from the base-case design (Sections 100 and 200).

By-product Markets

The economics of the base-case design and trade-off studies indicate that a better understanding and definition of the by-product markets is needed. The necessity of increasing total revenue via pentose and lignin by-product sale was discussed in Section 7. The uncertainties of the by-product values, saleability, and market size require the development of market data to define and/or establish by-product markets.

Technical Improvements

Once the feedstock and product market data have been clearly defined and the potential process, including projections of the virtues of technical improvements, are determined to be economically viable, then research and developed on these improvements should be considered. Areas of the base-case design requiring testing were outlined and discussed in Section 9.4, in addition to the need for vendor or pilot plant verification of equipment operability and reliability.

Research and development on potential process improvements are required for additional economic improvement in the base-case design. The areas identified in which potential improvements in the plant economics can be achieved are in increasing the efficiency of hydrolysis, increasing enzyme activity, and decreasing enzyme fermentation residence time.

The recovery of sodium hydroxide will give a \$0.15/gal reduction in the ethanol selling price. This alternate method reduces both chemical and disposal costs and should be included in any future design. The potential reduction in the required ethanol selling price for these individual research goals is secondary in importance to by-product sales, but the combined effect does enhance the attractiveness of the process.

TABLE 9.5-1

RESEARCH AND DEVELOPMENT PRIORITIES

1. Feedstock Research (what, how much, when, and market)

2. By-product Value

   A. Markets (current and potential)

   B. Technical Feasibility

      1. Ethanol from the C_5 fraction

      2. Furfural and acetic acid from the C_5 fraction

      3. Lignin derivations

3. Technical Improvements for the Enzyme Hydrolysis Process to be Proven Feasible

   A. Sodium Hydroxide Recovery

   B. Hydrolysis Yield, Residence Time, and Concentration Improvements

   C. Increase in Enzyme Activity

   D. Enzyme Residence Time Reduction

4. Less Viable Technical Improvements

   A. Lignin Removal after Hydrolysis

   B. Elimination of Evaporator

   C. Elimination of Chip Soak

9.6 CONCLUSIONS

In conclusion, the economic analysis of this study shows that significant revenue from pentose and lignin by-products must be obtained to allow this process to approach economic viability when viewed under the economic assumptions. Research is required to determine the most economical feedstock for producing ethanol that maximizes total revenue (products and by-products) and minimizes inhibitory compounds. Bench-scale data assumed in the base-case design must be verified under industrial-scale operating conditions to justify the base case.

The recovery of sodium hydroxide used to solubilize the lignin fraction of the steam exploded wood should be incorporated into future designs. Technical research and development in the areas of enzyme production and hydrolysis should be started after establishing the existence of by-product markets and the potential to achieve a competitive selling price for the production of ethanol from wood.

APPENDIX A

VENDOR LIST

The following is a list of vendors used in this study:

Aeroglide Corp., Raleigh, NC

ALFA-LAVAL Inc., Ft. Lee, NJ

Alpine American Corp., Natick, MA

Bacardi Corp., San Juan, PR

Bepex Corp., Minneapolis, MN

Bioengineering Association Inc., Newton, MA

Bird Machine Co., So. Walpole, MA

Carrier Corp., Syracuse, NY

Chishalm Corp., Cranston, RI

Dorr-Oliver Inc., Stamford, CT

Economics Laboratory Inc., Klenzade; St. Paul, MN

General Electric Co., Fitchburg, MA

IFE Systems Inc., Mahwah, NJ

Ingersoll-Rand Co., Implo Division, Nashua, NH

Joy Manufacturing, Denver Equipment Co., Englewood, CO

Kamyr, Inc., Glens Falls, NY

Kyowa Hakko Kogyo Co. Ltd., Tokyo, Japan

LSL Biolafitte Inc., Princeton, NJ

M.A. Olson Co. Inc., Topsfield, MA

Martin Engineering Co., Neponset, IL

Masonite Corp., Laurel, MI

Miles Laboratories, Clifton, NJ

Miller-Hofft, Richmond, VA

Modo-Chemetics, Vancouver, BC, Canada

New Brunswick Scientific Co., Edison, NJ

Niro Atomizer Inc., Columbia, MD

Novo Labortories Inc., Wilton, CT

Pennwalt Corp., Shaples Division, No. White Plains, NY

Rader Companies Inc., Portland, OR

Retel Inc., Atlanta, GA

Rosenblad Corp., Princeton, NJ

Solids Circulation Systems Inc., Boston, MA

Sunds Defribator, Inc., Minneapolis, MN

Tuthill Corp., Chicago, IL

United Technologies Elliot, Jeanette, PA

US Filter Corp., Chicago, IL

Williams Crusher Co., St. Louis, MO

Zimpro Inc., Rothschild, WI

APPENDIX B

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APPENDIX C

EQUIPMENT AND MOTOR LISTS

TABLE C-1

EQUIPMENT LIST

Item No.	Service	Comments
Section 100 - Pretreatment		
M-101A-F	Impregnation Vessels	
M-102	Sulfuric Acid Storage Tank	
P-101A-F	Acid Recycle Pumps	
P-102A-F	Sulfuric Acid Metering Pump	
V-105A-X	Air Cannons	
W-101	Inclined Conveyor Feeder	
W-102	Pretreatment Feed Conveyor	
W-103A,B	Impregnation Vessel Feed Conveyor	
W-104A-F	Impregnation Vessel Screw Discharge	
W-105A-F	Impregnation Product Screw Conveyor	
W-106	Central Product Belt Conveyor	
Section 200 - Steam Explosion/Wash		
M-201	MP Flash Vessel	
M-202	Vacuum Flash Vessel	
M-203A-D	Steam Explosion Feed Bins	
P-201A&B	Anaerobic Digester Feed Pump	1 operating, 1 spare
P-202A&B	Lignin Centrifuge Feed Pump	1 operating, 1 spare
P-203A&B	Water/Alkali Wash Feed Pump	1 operating, 1 spare
P-204A&B	Water Wash Recycle Pump	1 operating, 1 spare
R-201	Vacuum Pump	
T-201	Vacuum Flash Condenser	
V-201	Counter-Current Water/Alkali Wash	
V-203A-D	Steam Explosion Guns	
W-201	Vibrating Rotary Feeder	
W-202	Washer Screw Conveyor	
W-203	Washed Cellulose Lift Conveyor	
Section 300 - Enzyme Production		
G-301A-F	Fermenter Agitators	
G-302	Primary Seed Vessel Agitator	
G-303	Seed Culture Vessel Agitator	

Item No.	Service	Comments
M-301A,B	Enzyme Fermenter #1	
M-302A,B	Enzyme Fermenter #2	
M-303A,B	Enzyme Fermenter #3	
M-304	Ammonium Hydroxide Storage Tank	
M-305	Nutrient Storage Tank	
M-306	Primary Seed Fermenter	
M-307	Seed Culture Vessel	
P-301A,B	Fermenter No. 1 Recycle Pump	
P-302A,B	Fermenter No. 2 Recycle Pump	
P-303A,B	Fermenter No. 3 Product Pump	
P-304	Enzyme Seed Pump	
P-305	Seed Culture Pump	
P-308	Ammonia Pump	
P-309	CSL Pump	
R-301A,B	Air Compressor	
T-301A,B	Fermenter No. 1 Recycle Cooler	
T-302A,B	Fermenter No. 2 Recycle Cooler	
T-303A,B	Air Sparge Cooler	
T-304A,B	Air Compressor Intercooler	
V-301A,B	Air Cartridge Filter	
V-302	Clean In Place System	
W-302	Enzyme Fermenter Feed Conveyor	
Section 400 - Hydrolysis		
G-401A-L	Hydrolysis Reactor Agitators	
G-402A,B	Enzyme Recovery Tank Agitators	
G-403A,B	Hydrolysis Recycle Centrifuge	
G-404	Hydrolysis Centrifuge	
L-401A,B	Hydrolysis Reactors	
L-402A,B	Hydrolysis Reactors	
L-403A,B	Hydrolysis Reactors	
L-404A,B	Hydrolysis Reactors	
L-405A,B	Hydrolysis Reactors	
L-406A,B	Hydrolysis Reactors	
M-401A,B	Enzyme Recovery Tanks	
M-402	Sulfuric Acid Storage Tank	
P-401A&B	Enzyme Recovery Pump	1 operating, 1 spare
P-403A,B&C	Evaporator Feed Pump	2 operating, 1 spare

Item No.	Service	Comments
P-404A,B&C	Hydrolysis Recycle Pump	2 operating, 1 spare
P-405A,B	Sulfuric Acid Metering Pump	
P-407A,B	Hydrolysate Pumps	
T-401	Hydrolysis Dilution Cooler	
V-401	Counter-Current Wash	
W-401	Hydrolysis and Enzyme Feed Conveyor	
W-402A,B	Hydrolysis Reactor Feed Conveyor	
W-403	Water Wash Feed Conveyor	
W-404	Washer Screw Discharge	
W-405	Lignin Transfer Conveyor	
Section 500 - Evaporation		
M-501	Evaporator Feed Drum	
M-502	Disengagement Drum No. 1	
M-503	Disengagement Drum No. 2	
M-504	Disengagement Drum No. 3	
M-505	Disengagement Drum No. 4	
M-506	Disengagement Drum No. 5	
P-501	Evaporator Feed Pump	
P-502	Evaporator Circulation Pump No. 1	
P-503	Evaporator Circulation Pump No. 2	
P-504	Evaporator Circulation Pump No. 3	
P-505	Evaporator Circulation Pump No. 4	
P-506	Evaporator Circulation Pump No. 5	
P-507	Evaporator Condensate Pump No. 1	
P-508	Evaporator Condensate Pump No. 2	
P-509	Evaporator Condensate Pump No. 3	
R-501	Evaporator Vacuum Pump	
T-501A,B	Evaporator Surface Condenser	
T-502	Evaporator Chest No. 1	
T-503	Evaporator Chest No. 2	
T-504	Evaporator Chest No. 3	
T-505	Evaporator Chest No. 4	
T-506	Evaporator Chest No. 5	
T-509A-F	Evaporator Feed Heater	
Section 600 - Fermentation		
G-601	Mix Tank Agitator	
G-602	Immobilized Bead Tank Shower	

Item No.	Service	Comments
G-604A-C	1st Stage Bead Recovery Cyclone	
G-605A-C	2nd Stage Bead Recovery Cyclone	
L-601A-C	1st Stage Immobilized Bed Fermenter	
L-602A-C	2nd Stage Immobilized Bed Fermenter	
M-601	Yeast Hydration/Alginate Mix Tank	
M-602	Immobilized Bead Production Tank	
M-603A-C	1st Stage Vapor/Liquid Separator	
M-604A-C	2nd Stage Vapor/Liquid Separator	
P-601	Mix Tank Feed Pump	
P-602	Immobilized Bead Tank Feed Pump	
P-603	Beer Still Feed Pump	
P-604A,B&C	Refrigerated Water Circulation Pump	2 operating, 1 spare
P-605A-C&D	2nd Stage Fermenter Feed Pump	3 operating, 1 spare
R-601	Air Compressor	
R-602	Refrigerant Compressor	
T-601	Fermenter Feed Cooler	
T-602	Refrigeration Loop Cooler	
T-603	Refrigerant Condenser	
T-604	Fermenter Feed Chiller	
Section 700 - Distillation		
A-701	Beer Still	
A-702	Anhydrous Column	
A-703	Recovery Column	
M-701	Beer Still Reflux Drum	
M-702	Anhydrous System Decanter	
M-703	Fusel Oil Decanter	
M-704	Anhydrous Column Hold Tank	
M-705	Beer Still Feed Tank	
M-706	Degasser Drum	
M-707	Anhydrous Column Recycle Drum	
M-708	Anhydrous Column Reflux Drum	
M-709	Recovery Column Reflux Drum	
M-710	Entrainer Storage Tank	
M-713	Fusel Oil Storage Tank	
P-701A&B	Beer Still Reflux Pump	1 operating, 1 spare
P-702A&B	Anhydrous Column Reflux Pump	1 operating, 1 spare
P-703A&B	Recovery Column Reflux Pump	1 operating, 1 spare
P-704A&B	Beer Still Reboiler Circulation Pump	1 operating, 1 spare

Item No.	Service	Comments
P-705A&B	Ethanol Product Pump	1 operating, 1 spare
P-707A&B	Anhydrous Column Feed Pump	1 operating, 1 spare
P-708A&B	Backstillage Pump	1 operating, 1 spare
P-709	Anhydrous Column Rerun Pump	1 operating, 1 spare
P-712	Entrainer Makeup Pump	
P-713A&B	Recovery Column Bottoms Pump	1 operating, 1 spare
P-715	Anhydrous Column Recycle Pump	
P-716	Beer Still Pump	
P-721A&B	Anhydrous Column Reboiler Condensate Pump	1 operating, 1 spare
P-724	Fusel Oil Product Pump	
P-725A&B	Recovery Column Feed Pump	1 operating, 1 spare
T-701	Beer Still Feed Preheater	
T-702	Beer Still Trim Condenser	
T-703	Recovery Column Overhead Condenser	
T-704	Ethanol Product Cooler	
T-705	Evaporator Feed Preheater	
T-706	Beer Still Bottoms Cooler	
T-708	Beer Still Reboiler	
T-709	Recovery Column Reboiler	
T-711	Decanter Feed Cooler	
T-712	Fusel Oil Cooler	
T-713	Anhydrous Column Purge Cooler	
T-715	Anhydrous Column Overhead Condenser	
T-717	Anhydrous Column Hold Tank Feed Cooler	
T-719	Anhydrous Column Reboiler	
T-724	Anhydrous Column Hold Tank Vent Condenser	
T-726	Beer Still Vent Condenser	
T-727	Degasser Drum Vent Condenser	
Section 800 - Anaerobic Digestion		
G-801	Flare Stack	
L-801	Anaerobic Digester	
M-801A&B	Digester Feed Hold Tank	
M-802	Nutrient Storage Tank	
M-803	Gas Storage Sphere	
P-801	Digester Feed Pump	
P-802	Nutrient Feed Pump	
P-803	Digested Sludge Pump	
R-801	Methane-Rich Gas Compressor	

Item No.	Service	Comments
T-801	Digester Feed Cooler	
Section 900 - Boiler		
B-901	Boiler	
G-901	Multiclone	
G-902	Baghouse	
G-903	Stack	
G-904	Lignin Centrifuge	
P-901	Fuel Oil Unloading Pump	
P-902	Fuel Oil Pump	
Q-901	Lignin Day Bin	
Q-902	Ash Silo	
Q-904	No. 2 Fuel Oil Storage Tank	
R-901	Primary Air Fan	
R-902	Induced Draft Fan	
T-901	Primary Economizer	
T-902	Superheater	
T-903	Air Preheater	
W-901	Boiler Screw Feeder	
W-902	Ash Silo Feed Conveyor	
W-903	Lignin Day Bin Screw Feeder	
W-904	Boiler Wood Chip Conveyor	
W-905	Boiler Wood Chip Screw Feeder	
Section 1000 - Wood Handling		
G-1001	Truck Scale	
G-1002	Front End Loader	
G-1003	Truck-Trailer Dumper	
G-1004	Truck Receiving Yard Pit	
G-1005	Scalping Screen	
G-1006	Primary Magnetic Separator	
G-1007	Secondary Magnetic Separator	
G-1008	Stone Trap	
G-1009	Oversized Wood Chipper	
G-1010A,B	Three-Deck Chip Screens	
G-1011	Belt Conveyor Scale	
G-1012	Chip Storage Silo	
G-1013	Vibrating Screen	

Item No.	Service	Comments
Q-1001	Truck Receiving Hopper	
Q-1002	Surge Bin	
W-1001	Unloading Bin Drag Chain	
W-1002	Scalping Screen Feed Drag Chain	
W-1003	Double Wing Belt Stacker	
W-1004	Stacker Feed Conveyor	
W-1005A,B	Traveling Scraper/Reclaimer	
W-1006	Storage Pile Transfer Conveyor	
W-1007A,B	Reclaimer Belt Conveyor	
W-1008	Wood Chip Elevating Conveyor	
W-1009	Fines Transfer Conveyor	
W-1010	Large Chip Transfer Conveyor	
W-1011	Medium Chip Transfer Conveyor	
W-1012	Cleaned Wood Chip Transfer Conveyor	
W-1013	Boiler Fuel Transfer Conveyor	
Section 1100 - Cooling Water		
G-1101	Cooling Tower	
G-1102A-G	Induced Draft Fans	
P-1101A,B&C	Cooling Water Circulating Pumps	2 operating, 1 spare
V-1101	Inhibitor Feed System	
V-1102	Acid Feed System	
V-1103	Chlorination System	
Section 1200 - Waste Treatment/Vent Scrubbing		
A-1201	CO_2 Wash Column	
A-1202	Vent System Scrubber	
G-1201	Oil/Water Separator	
G-1202	Primary Clarifier	
G-1203	Sludge Thickener	
G-1204A,B&C	Belt Filter Press	2 operating, 1 spare
G-1205A,B&C	Trickling Filter	2 operating 1 spare
G-1206	Secondary Clarifier	
G-1207	Final Clarifier	
G-1209	Neutralization Tank Agitator	
G-1210	Primary Clarifier Rake	
G-1211	Secondary Clarifier Rake	
G-1212A-H	Aeration Basin Agitator	
G-1213	Final Clarifier Rake	
G-1214	Sludge Thickener Rake	
G-1215	Polymer Feed Tank Agitator	

Item No.	Service	Comments
G-1216	Sludge Mixing Tank Agitator	
G-1217A,B&C	Trickling Filter Fans	2 operating, 1 spare
M-1201	Settling Basin	
M-1202	Sulfuric Acid Storage	
M-1203	Sodium Hydroxide Storage	
M-1204	Ammonium Hydroxide Storage	
M-1205	Phosphoric Acid Storage	
M-1206	Equalization Basin	
M-1207	Neutralization Tank	
M-1208	Polymer Storage Tank	
M-1209	Polymer Feed Tank	
M-1210	Sludge Mixing Tank	
M-1211	Filtrate Collection Tank	
M-1212	Distribution Sump	
M-1213	Aeration Basin	
M-1214	Discharge Monitoring Sump	
P-1201A&B	Settling Basin Effluent Pump	1 operating, 1 spare
P-1202A&B	Separator Water Pump	1 operating, 1 spare
P-1203A&B	Sulfuric Acid Metering Pump	1 operating, 1 spare
P-1204A&B	Sodium Hydroxide Metering Pump	1 operating, 1 spare
P-1205A&B	Primary Clarifier Sludge Pump	1 operating, 1 spare
P-1206	Polymer Transfer Pump	
P-1207A&B	Polymer Feed Pump	1 operating, 1 spare
P-1208A&B	Filtrate Transfer Pump	1 operating, 1 spare
P-1209A,B&C	Trickling Filter Feed Pump	2 operating, 1 spare
P-1210A&B	Secondary Clarifier Sludge Pump	1 operating, 1 spare
P-1211A&B	Final Clarifier Sludge Pump	1 operating, 1 spare
P-1212A&B	Neutralization Feed Pump	1 operating, 1 spare
P-1213A&B	Neutralized Effluent Pump	1 operating, 1 spare
P-1214A&B	Ammonium Hydroxide Metering Pump	1 operating, 1 spare
P-1215A&B	Phosophoric Acid Metering Pump	1 operating, 1 spare
P-1216A&B	Clean Water Discharge Pump	1 operating, 1 spare
P-1217A&B	Belt Filter Washwater Pump	1 operating, 1 spare
P-1218	CO_2 Wash Column Pump	
P-1219	Vent Scrubber Pump	
P-1220A&B	Sludge Thickener Underflow Pump	1 operating, 1 spare
P-1221A,B&C	Sludge Transfer Pump	2 operating, 1 spare
R-1201	CO_2 Wash Column Blower	
R-1202	Vent Scrubber Blower	
W-1201	Thickened Sludge Conveyor	
V-1201	Chlorination System	

Item No.	Service	Comments
Section 1300 - Chemical Handling		
M-1301	Sodium Hydroxide Storage Tank	
M-1302	Sulfuric Acid Storage Tank	
P-1301	Sodium Hydroxide Feed Pump	
P-1302	Sulfuric Acid Feed Pump	
Section 1400 - Product Storage and Unloading		
G-1401A,B	Alcohol Truck Loading Station	
P-1401A,B	Alcohol Loading Pump	
P-1402	Gasoline Unloading Pump	
P-1403	Gasoline Metering Pump	
Q-1401	Denatured Alcohol Storage Tank	
Q-1402	Gasoline Storage Tank	
Section 1500 - Water Treatment/Condensate Return		
M-1501	Caustic Storage Tank	
M-1502	Hydrochloric Acid Storage Tank	
M-1503A,B	Cation Vessels	
M-1504A,B	Anion Vessels	
M-1505A,B	Mixed Bed Vessels	
M-1506	Demineralized Water Tank	
M-1507	Deaerator	
P-1501	Caustic Pump	
P-1502	Hydrochloric Acid Pump	
P-1503	Deaerator Feed Pump	
P-1504	Boiler Feedwater Pump	
Section 1600 - Instrument Air/Fire Protection		
G-1601	Halon Fire Extinguishing System	
G-1602	Foam System	
M-1601	Instrument Air Receiver	
M-1602	Instrument/Service Air Receiver	
P-1601	Motor-Driven Fire Pump	
P-1602	Jockey Fire Pump	
P-1603	Diesel-Driven Fire Pump	
P-1604A&B	Diesel Fuel Oil Pumps	1 operating, 1 spare

Item No.	Service	Comments
Q-1601	Fire Protection Water Storage	
Q-1602	Hydropneumatic Tank	
Q-1603	Diesel Fuel Oil Tank	
R-1601	Instrument Air Compressor	
R-1602	Service Air Compressor	
V-1603A,B	Air Prefilters	
V-1604A,B	Air Afterfilters	
V-1607A,B	Air Dryers	

TABLE C-2

SIZED EQUIPMENT LIST

General Equipment Summary

Item No.	Service	Rated Hp (each)	Size (each)	Comments
Section 100 - Pretreatment				
V-105A-X	Air Cannons			Vendor Package
Section 200 - Steam Explosion/Wash				
V-201	Counter Current Water/Alkali Wash	150	840 ft^2	Vendor Package
V-203A-D	Steam Explosion Guns		26" dia	Vendor Package
Section 300 - Enzyme Production				
G-301A-F	Fermenter Agitators	150		Vendor Package
G-302	Primary Seed Vessel Agitator	3		Vendor Package
G-303	Seed Culture Vessel Agitator	10		Vendor Package
V-301A,B	Air Cartridge Filter		24" dia	Vendor Package
V-302	Clean in Place System	150		Vendor Package
Section 400 - Hydrolysis				
G-401A-L	Hydrolysis Reactor Agitators	100		Vendor Package
G-402A,B	Enzyme Recovery Tank Agitators	125		Vendor Package
G-403A,B	Hydrolysis Recycle Centrifuge	500		Vendor Package
G-404	Hydrolysis Centrifuge	350		Vendor Package
L-401A,B	Hydrolysis Reactors		35,400 ft^3	
L-402A,B	Hydrolysis Reactors		35,400 ft^3	
L-403A,B	Hydrolysis Reactors		35,400 ft^3	
L-404A,B	Hydrolysis Reactors		35,400 ft^3	
L-405A,B	Hydrolysis Reactors		35,400 ft^3	
L-406A,B	Hydrolysis Reactors		35,400 ft^3	
V-401	Counter-Current Wash	300		Vendor Package
Section 600 - Fermentation				
G-601	Mix Tank Agitator			Vendor Package
G-602	Immobilized Bead Tank Shower			Vendor Package
G-604A-C	1st Stage Bead Recovery Cyclone			Vendor Package
G-605A-C	2nd Stage Bead Recovery Cyclone			Vendor Package
L-601A-C	1st Stage Immobilized Bed Fermenter			Vendor Package

General Equipment Summary

Item No.	Service	Rated Hp (each)	Size (each)	Comments
L-602A-C	2nd Stage Immobilized Bed Fermenter			Vendor Package
Section 800 - Anaerobic Digestion				
L-801	Anaerobic Digester			Vendor Package
G-801	Flare Stack			Vendor Package
Section 900 - Boiler				
B-901	Boiler			Vendor Package
G-901	Multiclone			Vendor Package
G-902	Baghouse			Vendor Package
G-903	Stack			
G-904	Lignin Centrifuge	300		Vendor Package
Section 1000 - Wood Feeding				
G-1001	Truck Scale		50 ton	
G-1002	Front End Loader		3 yd^3	
G-1003	Truck-Trailer Dumper		50 ton	
G-1004	Truck Receiving Yard Pit			
G-1005	Scalping Screen		1 5/8 in. screen	Vendor Package
G-1006	Primary Magnetic Separator		24 in. wide	
G-1007	Secondary Magnetic Separator		24 in. wide	
G-1008	Stone Trap		24 in. wide	
G-1009	Oversized Wood Chipper	60		
G-1010A,B	Three-Deck Chip Screens		14,400 ft^2	
G-1011	Belt Conveyor Scale		24 in. wide	
G-1012	Chip Storage Silo	25	9,600 ft^3	W/Screw Reclaimer
G-1013	Vibrating Screen			Vendor Package
Section 1100 - Cooling Water				
G-1101	Cooling Tower		13,000 gpm	Vendor Package
G-1102A-G	Induced Draft Fans	60		
V-1101	Inhibitor Feed System	2.5		Vendor Package
V-1102	Acid Feed System	2.5		Vendor Package
V-1103	Chlorination System			Vendor Package

General Equipment Summary

Item No.	Service	Rated Hp (each)	Size (each)	Comments
Section 1200 - Waste Treatment/Vent Scrubbing				
G-1201	Oil/Water Separator			Vendor Package
G-1202	Primary Clarifier		32' dia	Vendor Package
G-1203	Sludge Thickener		70' dia	Vendor Package
G-1204A,B&C	Belt Filter Press	5		Vendor Package
G-1205A,B&C	Trickling Filter		45' dia x 26'	Vendor Package
G-1206	Secondary Clarifier		33' dia	Vendor Package
G-1207	Final Clarifier		33' dia	Vendor Package
G-1209	Neutralization Tank Agitator	0.5		Vendor Package
G-1210	Primary Clarifier Rake	0.5		Vendor Package
G-1211	Secondary Clarifier Rake	0.5		Vendor Package
G-1212A-H	Aeration Basin Agitator	60		Vendor Package
G-1213	Final Clarifier Rake	0.5		Vendor Package
G-1214	Sludge Thickener Rake	0.5		Vendor Package
G-1215	Polymer Feed Tank Agitator	1		Vendor Package
G-1216	Sludge Mixing Tank Agitator	2		Vendor Package
G-1217A,B&C	Trickling Filter Fans	1		Vendor Package
Section 1400 - Product Storage and Unloading				
G-1401A,B	Alcohol Truck Loading Station			Vendor Package
Section 1600 - Instrument Air/Fire Protection				
G-1601	Halon Fire Extinguishing System			Vendor Package
G-1602	Foam System		5,000 gal	Vendor Package
V-1603A,B	Air Prefilters			
V-1604A,B	Air Afterfilters			
V-1607A,B	Air Dryers			Dessicant

TABLE C-2 (Cont)

Pump Summary

Item No.	Service	Liquid Handled	Pump Type	Capacity (each) (gpm)	ΔP (psi)	Driver Type	Rated hp (each)	Material	Comments
Section 100 - Pretreatment									
P-101A-F	Acid Recycle Pumps	Dilute Acid	Cent	1,250	70	Motor	75	SS	
P-102A-F	Sulfuric Acid Metering Pump	H_2SO_4	Metering	5	25	Motor	0.5	CS	
Section 200 - Steam Explosion/Wash									
P-201A&B	Anaerobic Digester Feed Pump	H_2O	Cent	340	38	Motor	15	CS	
P-202A&B	Lignin Centrifuge Feed Pump	Lignin Slurry	Cent	311	25	Motor	7.5	CS	
P-203A&B	Water Alkali Wash Feed Pump	10% Solids	Cent	739	25	Motor	15	SS	
P-204A&B	Water Wash Recycle Pump	H_2O	Cent	510	25	Motor	10	CS	
Section 300 - Enzyme Production									
P-301A,B	Fermenter No. 1 Recycle Pump	Thick Slurry	Cent	310	28	Motor	10	CS	
P-302A,B	Fermenter No. 2 Recycle Pump	Thick Slurry	Cent	400	28	Motor	10	CS	
P-303A,B	Fermenter No. 3 Product Pump	Thick Slurry	Cent	400	28	Motor	10	CS	
P-304	Enzyme Seed Pump	Slurry	Cent	60	55	Motor	3	CS	
P-305	Seed Culture Pump	Slurry	Cent	5	55	Motor	0.5	CS	
P-308	Ammonia Pump	NH_3/H_2O	Metering	5-8	25	Motor	0.5	CS	
P-309	CSL Pump	Corn Steep Liquor	Cent	15	25	Motor	0.5	CS	
Section 400 - Hydrolysis									
P-401A&B	Enzyme Recovery Pump	Sugar Solution	Cent	330	30	Motor	10	CS	
P-403A,B&C	Evaporator Feed Pump	Sugar Solution	Cent	500	25	Motor	15	CS	
P-404A,B&C	Hydrolysis Recycle Pump	Thick Slurry	Cent	170	25	Motor	5	CS	
P-405A,B	Sulfuric Acid Metering Pump	H_2SO_4	Metering	0-5	25	Motor	0.5	CS	
P-407A,B	Hydrolysate Pumps	Sugar Solution	Cent	1,250	28	Motor	30	CS	
Section 500 - Evaporation									
P-501	Evaporator Feed Pump	Slurry	Cent			Motor		SS	Vendor Package
P-502	Evaporator Circulation Pump No. 1	Slurry	Cent			Motor		SS	Vendor Package
P-503	Evaporator Circulation	Slurry	Cent			Motor		SS	Vendor Package

Pump Summary

Item No.	Service	Liquid Handled	Pump Type	Capacity (each) (gpm)	ΔP (psi)	Driver Type	Rated hp (each)	Material	Comments
	Pump No. 2								
P-504	Evaporator Circulation Pump No. 3	Slurry	Cent			Motor		SS	Vendor Package
P-505	Evaporator Circulation Pump No. 4	Slurry	Cent			Motor		SS	Vendor Package
P-506	Evaporator Circulation Pump No. 5	Slurry	Cent			Motor		SS	Vendor Package
P-507	Evaporator Condensate Pump No. 1	H_2O	Cent			Motor		SS	Vendor Package
P-508	Evaporator Condensate Pump No. 2	H_2O	Cent			Motor		SS	Vendor Package
P-509	Evaporator Condensate Pump No. 3	H_2O	Cent			Motor		SS	Vendor Package
Section 600 - Fermentation									
P-601	Mix Tank Feed Pump	H_2O	Cent			Motor			Vendor Package
P-602	Immobilized Bead Tank Feed Pump	H_2O	Cent			Motor			Vendor Package
P-603	Beer Still Feed Pump	H_2O/ETOH	Cent	313	5	Motor	10	CS	
P-604A,B&C	Refrigerated Water Circulation Pump	H_2O	Cent	713	7	Motor	4	CS	Vendor Package
P-605A-C&D	2nd Stage Fermenter Feed Pump	Broth	Cent			Motor			Vendor Package
Section 700 - Distillation									
P-701A&B	Beer Still Reflux Pump	ETOH/H_2O/ Glucose/ Organic Acids	Cent	160	53	Motor	7.5	CS	
P-702A&B	Anhydrous Column Reflux Pump	ETOH/ Cyclohexane/ H_2O/MEOH	Cent	200	56	Motor	10	CS	
P-703A&B	Recovery Column Reflux Pump	ETOH/ Cyclohexane/ H_2O/MEOH	Cent	60	46	Motor	3	CS	
P-704A&B	Beer Still Reboiler Circulation Pump	ETOH/H_2O/ Glucose/ Organic Acids	Cent	300	6	Motor	1.5	CS	
P-705A&B	Ethanol Product Pump	ETOH/H_2O	Cent	40	24	Motor	1.5	CS	
P-707A&B	Anhydrous Column Feed Pump	ETOH/H_2O/ MEOH	Cent	85	56	Motor	5	CS	
P-708A&B	Backstillage Pump	H_2O/Solids	Cent	390	30	Motor	10	CS	

TABLE C-2 (Cont)

Pump Summary

Item No.	Service	Liquid Handled	Pump Type	Capacity (each) (gpm)	ΔP (psi)	Driver Type	Rated hp (each)	Material	Comments
P-709	Anhydrous Column Rerun Pump	ETOH/H_2O/MEOH	Cent	10	56	Motor	0.5	CS	
P-712	Entrainer Makeup Pump	Cyclohexane	Cent	2	20	Motor	0.5	CS	
P-713A&B	Recovery Column Bottoms Pump	Dirty H_2O	Cent	3	38	Motor	0.5	CS	
P-715	Anhydrous Column Recycle Pump	ETOH/H_2O/MEOH/Cyclohexane	Cent	45	61	Motor	3	CS.	
P-716	Beer Still Feed Pump	H_2O/ETOH/Solids	Cent	370	68	Motor	20	CS	
P-721A&B	Anhydrous Column Reboiler Condensate Pump	Dirty H_2O	Cent	25	59	Motor	1.5	CS	
P-724	Fusel Oil Product Pump	Fusel Oils	Cent	2	25	Motor	0.5	CS	
P-725A&B	Recovery Column Feed Pump	ETOH/Cyclohexane/H_2O/MEOH	Cent	30	53	Motor	1.5	CS	
Section 800 - Anaerobic Digestion									
P-801	Digester Feed Pump	H_2O/sugar	Cent	317	24	Motor	10	CS	
P-802	Nutrient Feed Pump	Nutrient Solution	Cent				0.5	SS	Vendor Package
P-803	Digested Sludge Pump	Sludge	Cent				10	CS	Vendor Package
Section 900 - Boiler									
P-901	Fuel Oil Unloading Pump	Fuel Oil	Cent	200	30	Motor	7.5	CS	
P-902	Fuel Oil Pump	Fuel Oil	Cent	45	23	Motor	2	CS	
Section 1100 - Cooling Water									
P-1101A,B&C	Cooling Water Circulating Pumps	H_2O	Cent	6,500	43	Motor	250	CS	
Section 1200 - Waste Treatment/Vent Scrubbing									
P-1201A&B	Settling Basin Effluent Pump	H_2O	Cent			Motor	3	CS	Vertical Shaft
P-1202A&B	Separator Water Pump	H_2O	Cent			Motor	3	CS	Vertical Shaft
P-1203A&B	Sulfuric Acid Metering Pump	H_2SO_4/H_2O	Metering			Motor	1	CS	
P-1204A&B	Sodium Hydroxide Metering Pump	NaOH/H_2O	Metering			Motor	1	CS	

TABLE C-2 (Cont)

Pump Summary

Item No.	Service	Liquid Handled	Pump Type	Capacity (each) (gpm)	ΔP (psi)	Driver Type	Rated hp (each)	Material	Comments
P-1205A&B	Primary Clarifier Sludge Pump	Sludge	Cent			Motor	5	CS	
P-1206	Polymer Transfer Pump	Polymer	Cent			Motor	1	CS	
P-1207A&B	Polymer Feed Pump	Polymer	Cent			Motor	1	CS	
P-1208A&B	Filtrate Transfer Pump	H_2O	Cent			Motor	5	CS	
P-1209A,B&C	Trickling Filter Feed Pump	H_2O	Cent			Motor	70	CS	
P-1210A&B	Secondary Clarifier Sludge Pump	Sludge	Cent			Motor	5	CS	
P-1211A&B	Final Clarifier Sludge Pump	Sludge	Cent			Motor	5	CS	
P-1212A&B	Neutralization Feed Pump	H_2O	Cent			Motor	2	CS	Vertical Shaft
P-1213A&B	Neutralized Effluent Pump	H_2O	Cent			Motor	3	CS	Vertical Shaft
P-1214A&B	Ammonium Hydroxide Metering Pump	NH_4OH/ H_2O	Metering			Motor	1	CS	
P-1215A&B	Phosophoric Acid Metering Pump	H_3PO_4/ H_2O	Metering			Motor	1	CS	
P-1216A&B	Clean Water Discharge Pump	H_2O	Cent			Motor	25	CS	Vertical Shaft
P-1217A&B	Belt Filter Washwater Pump	H_2O	Cent			Motor	3	CS	Vertical Shaft
P-1218	CO_2 Wash Column Pump	ETOH/CO_2	Cent	11.7	9	Motor	0.75	CS	
P-1219	Vent Scrubber Pump	ETOH/H_2O/ MEOH	Cent	1.5	9	Motor	0.5	CS	
P-1220A&B	Sludge Thickener Underflow Pump	Sludge	Positive Displacement			Motor	3	CS	
P-1221A,B&C	Sludge Transfer Pump	Sludge	Positive Displacement			Motor	2	CS	
Section 1300 - Chemical Handling									
P-1301	Sodium Hydroxide Feed Pump	NaOH/H_2O	Cent	5	62	Motor	1	CS	
P-1302	Sulfuric Acid Feed Pump	H_2SO_4	Cent	2	78	Motor	0.5	CS	
Section 1400 - Product Storage and Handling									
P-1401A,B	Denatured Alcohol Loading Pump	ETOH/ Gasoline	Cent	250	385	Motor	15	CS	
P-1402	Gasoline Unloading Pump	Gasoline	Cent	200	33	Motor	10	CS	
P-1403	Gasoline Metering Pump	Gasoline	Metering	2	30	Motor	0.5	CS	
Section 1500 - Water Treatment/Condensate Return									
P-1501	Caustic Pump	NaOH/H_2O	Cent			Motor	0.5		Vendor Package
P-1502	Hydrochloric Acid Pump	HCl/H_2O	Cent			Motor	0.5		Vendor Package
P-1503	Deaerator Feed Pump	H_2O	Cent	85	34	Motor	5	CS	
P-1504	Boiler Feedwater Pump	H_2O	Cent	280	761	Motor	250	CS	

TABLE C-2 (Cont)

Pump Summary

Item No.	Service	Liquid Handled	Pump Type	Capacity (each) (gpm)	ΔP (psi)	Driver Type	Rated hp (each)	Material	Comments
Section 1600 - Instrument Air/Fire Protection									
P-1601	Motor-Driven Fire Pump	H_2O	Cent	2,500	125	Motor	300	CS	Normally Off
P-1602	Jockey Fire Pump	H_2O	Cent	300	25	Motor	10	CS	
P-1603	Diesel-Driven Fire Pump	H_2O	Cent	2,500	120	Diesel		CS	Normally Off
P-1604A&B	Diesel Fuel Oil Pumps	Diesel Fuel	Cent	100	40	Motor	5	CS	Normally Off

TABLE C-2 (Cont)

Compressor/Blower summary

Item No.	Service	Vapor Handled	Capacity (ACFM Suction) (each)	Equipment Type	Driver Type	Rated Hp (each)	Material	Comments
Section 200 - Steam Explosion/Wash								
R-201	Vacuum Pump	Vacuum Flash Vapor	8	Rotary	Motor	1	CS	5 psia Vacuum
Section 300 - Enzyme Production								
R-301A,B	Air Compressor	Air	16,000	Rotary	Steam Turbine	2277	CS	
Section 500 - Evaporation								
R-501	Evaporator Vacuum Pump	Mixture		Rotary	Motor		CS	Vendor Package
Section 600 - Fermentation								
R-601	Air Compressor	Air		Rotary	Motor	10	CS	Vendor Package
R-602	Refrigerant Compressor	Refrigerant		Rotary	Motor	1233	CS	Vendor Package
Section 800 - Anaerobic Digestion								
R-801	Methane-Rich Gas Compressor	CH_4/CO_2		Rotary	Motor	175	CS	Vendor Package
Section 900 - Boiler								
R-901	Primary Air Fan	Air		Rotary	Motor	1410	CS	Vendor Package
R-902	Induced Draft Fan	Air		Rotary	Motor	1410	CS	Vendor Package
Section 1200 - Waste Treatment/Vent Scrubbing								
R-1201	CO_2 Wash Column Blower	CO_2	1771	Rotary	Motor	150	CS	
R-1202	Vent Scrubber Blower	Mixture	117	Rotary	Motor	15	CS	
Section 1600 - Instrument Air/Fire Protection								
R-1601	Instrument Air Compressor	Air	600	Rotary	Motor	125	CS	
R-1602	Service Air Compressor	Air	600	Rotary	Motor	125	CS	

TABLE C-2 (Cont)

Conveyor Summary

Item No.	Service	Solids Handled	Equipment Type	Size (in.)	Length (ft)	Rated Drive Hp (each)	Material	Comments
Section 100 - Pretreatment								
W-101	Inclined Conveyor Feeder	Wood Chips	Belt	24	135	3	CS	Enclosed
W-102	Pretreatment Feed Conveyor	Wood Chips	Belt	24	37	3	CS	Enclosed
W-103A,B	Impregnation Vessel Feed Conveyor	Wood Chips	Belt	24	70	5	CS	Enclosed
W-104A-F	Impregnation Vessel Screw Discharge	Wood Chips	Screw	8	10	50	SS	10 Parallel Screws
W-105A-F	Impregnation Product Screw Conveyor	Wood Chips	Screw	24	24	10	SS	
W-106	Central Product Belt Conveyor	Wood Chips	Belt	18	325	12.5	SS	Enclosed
Section 200 - Steam Explosion/Wash								
W-201	Vibrating Rotary Feeder	Wood Chips			10	5	SS	
W-202	Washer Screw Conveyor	Exploded Wood	Screw	16	26	7.5	SS	Enclosed
W-203	Washed Cellulose Lift Conveyor	Exploded Wood	Belt	18	317	12.5	SS	Enclosed
Section 300 - Enzyme Production								
W-302	Enzyme Fermenter Feed Conveyor	Exploded Wood	Drag Chain	13	160	30	CS	Enclosed Loop Conveyor
Section 400 - Hydrolysis								
W-401	Hydrolysis and Enzyme Feed Conveyor	Exploded Wood	Drag Chain	13	240	40	CS	Enclosed
W-402A,B	Hydrolysis Reactor Feed Conveyor	Exploded Wood	Drag Chain	13	280	50	CS	Enclosed
W-403	Water Wash Feed Conveyor	Exploded Wood	Screw	12	20	3	SS	
W-404	Washer Screw Discharge	Exploded Wood	Screw	12	20	3	SS	
W-405	Lignin Transfer Conveyor	Exploded Wood	Belt	18	52	4	CS	Enclosed
Section 900 - Boiler								
W-901	Boiler Screw Feeder	Lignin/ Exploded Wood	Screw	14	10	3	CS	
W-902	Ash Silo Feed Conveyor	Ash	Screw	9	20	1.5	CS	
W-903	Lignin Day Bin Screw Feeder	Lignin	Screw	12	10	1	CS	

Conveyor Summary

Item No.	Service	Solids Handled	Equipment Type	Size (in.)	Length (ft)	Rated Drive Hp (each)	Material	Comments
W-904	Boiler Wood Chip Conveyor	Wood Chips	Belt	18	280	25	CS	Enclosed
W-905	Boiler Wood Chip Screw Feeder	Wood Chips	Screw	12	10	1	CS	
Section 1000 - Wood Feeding								
W-1001	Unloading Bin Drag Chain	Wood Chips	Drag Chain	24	70	75	CS	
W-1002	Scalping Screen Feed Drag Chain	Wood Chips	Drag Chain	24	80	75	CS	
W-1003	Double Wing Belt Stacker	Wood Chips	Belt	36	50	20	CS	Traveling
W-1004	Stacker Feed Conveyor	Wood Chips	Belt	36	550	40	CS	
W-1005A,B	Traveling Scraper/Reclaimer	Wood Chips	Belt		90	130	CS	45° Angle
W-1006	Storage Pile Transfer Conveyor	Wood Chips	Belt	24	250	25	CS	
W-1007A,B	Reclaimer Belt Conveyor	Wood Chips	Belt	24	550	30	CS	
W-1008	Wood Chip Elevating Conveyor	Wood Chips	Belt	24	250	25	CS	
W-1009	Fines Transfer Conveyor	Wood Chips	Belt	18	30	2	CS	
W-1010	Large Chip Transfer Conveyor	Wood Chips	Belt	18	30	2	CS	
W-1011	Medium Chip Transfer Conveyor	Wood Chips	Belt	24	30	3	CS	
W-1012	Cleaned Wood Chip Transfer Conveyor	Wood Chips	Belt	24	230	25	CS	
W-1013	Boiler Fuel Transfer Conveyor	Wood Chips	Belt	18	250	20	CS	
Section 1200 - Waste Treatment/Vent Scrubbing								
W-1201	Thickened Sludge Conveyor	Wet Sludge	Belt			2	CS	Vendor Package

TABLE C-2 (Cont)

Drum Summary

Item No.	Service	Diameter (ft)	Height (ft)	Horiz/ Vert	Material	Comments
Section 100 - Pretreatment						
M-101A-F	Impregnation Vessels	35	48	V	CS/FRP Lining	Live Bottom Bin
M-102	Sulfuric Acid Storage Tank	3.5	14	V	CS	
Section 200 - Steam Explosion/Wash						
M-201	MP Flash Vessel	12'6"	29	V	SS	
M-202	Vacuum Flash Vessel	10	10	V	SS	
M-203 A-D	Steam Explosion Feed Bins	9	14.5	V (Cone)	CS/Epoxy Lining	Tapers to 2' Diameter
Section 300 - Enzyme Production						
M-301A,B	Enzyme Fermenter No. 1	18	60	V	CS/Epoxy Lining	
M-302A,B	Enzyme Fermenter No. 2	18	60	V	CS/Epoxy Lining	
M-302A,B	Enzyme Fermenter No. 3	18	60	V	CS/Epoxy Lining	
M-304	Ammonium Hydroxide Storage Tank	8.5	39	H	CS	
M-305	Nutrient Storage Tank	15	15	V	CS	
M-306	Primary Seed Fermenter	4	8	V	304SS	Vendor Package
M-307	Seed Culture Vessel	1	2	V	304SS	Vendor Package
Section 400 - Hydrolysis						
M-401A,B	Enzyme Recovery Tanks	40	88	V	CS/Epoxy Lining	
M-402	Sulfuric Acid Storage Tank	3.5	14	V	CS	
Section 500 - Evaporation						
M-501	Evaporator Feed Drum					Vendor Package
M-502	Disengagement Drum No. 1					Vendor Package
M-503	Disengagement Drum No. 2					Vendor Package
M-504	Disengagement Drum No. 3					Vendor Package
M-505	Disengagement Drum No. 4					Vendor Package
M-506	Disengagement Drum No. 5					Vendor Package

Drum Summary

Item No.	Service	Diameter (ft)	Height (ft)	Horiz/ Vert	Material	Comments
Section 600 - Fermentation						
M-601	Yeast Hydration/Alginate Mix Tank					Vendor Package
M-602	Immobilized Bead Production Tank					Vendor Package
M-603A-C	1st Stage Vapor/Liquid Separator					Vendor Package
M-604A-C	2nd Stage Vapor/Liquid Separator					Vendor Package
Section 700 - Distillation						
M-701	Beer Still Reflux Drum	5	10	H	CS	
M-702	Anhydrous System Decanter	12.5	18	V	CS	
M-703	Fusel Oil Decanter	3	6	V	CS	
M-704	Anhydrous Column Hold Tank	20	22	V	CS	
M-705	Beer Still Feed Tank	9	16	V	CS	
M-706	Degasser Drum	1.5	4	V	CS	
M-707	Anhydrous Column Recycle Drum	3.5	7	V	CS	
M-708	Anhydrous Column Reflux Drum	4	12	H	CS	
M-709	Recovery Column Reflux Drum	3	9	H	CS	
M-710	Entrainer Storage Tank	4	8	H	CS	
M-713	Fusel Oil Storage Tank	3	4	H	CS	
Section 800 - Anaerobic Digestion						
M-801	Digester Feed Hold Tank					Vendor Package
M-802	Nutrient Storage Tank					Vendor Package
M-803	Gas Storage Sphere	30.25		Spherical	CS	100 psig Working Pressure
Section 1200 - Waste Treatment/Vent Scrubbing						
M-1201	Settling Basin			H	Concrete	76,200 ft^3 Basin
M-1202	Sulfuric Acid Storage	6	25	H	CS	
M-1203	Sodium Hydroxide Storage	6	25	H	CS	
M-1204	Ammonium Hydroxide Storage	6	25	H	CS	
M-1205	Phosphoric Acid Storage	6	25	H	CS	
M-1206	Equalization Basin			H	Concrete	10,000 ft^3 Basin
M-1207	Neutralization Tank			H	CS	400 ft^3
M-1208	Polymer Storage Tank			V		Vendor Package

TABLE C-2 (Cont)

Drum Summary

Item No.	Service	Diameter (ft)	Height (ft)	Horiz/ Vert	Material	Comments
M-1209	Polymer Feed Tank			V		Vendor Package
M-1210	Sludge Mixing Tank			V		Vendor Package
M-1211	Filtrate Collection Tank			H		Vendor Package
M-1212	Distribution Sump			H	Concrete	8,000 ft³ Basin
M-1213	Aeration Basin			H	Concrete	64,000 ft³ Basin
M-1214	Discharge Monitoring Sump			H	Concrete	2,000 ft³ Basin
Section 1300 - Chemical Handling						
M-1301	Sodium Hydroxide Storage Tank	25	30	H	CS	
M-1302	Sulfuric Acid Storage Tank	20	24	H	CS	
Section 1500 - Water Treatment/Condensate Return						
M-1501	Caustic Storage Tank	6	25	H	CS	
M-1502	Hydrochloric Acid Storage Tank	6	25	H	CS	
M-1503A,B	Cation Vessels					Vendor Package
M-1504A,B	Anion Vessels					Vendor Package
M-1505A,B	Mixed Bed Vessels					Vendor Package
M-1506	Demineralized Water Tank					Vendor Package
M-1507	Deaerator					Vendor Package
Section 1600 - Instrument Air/Fire Protection						
M-1601	Instrument Air Receiver	3.5	10	V	CS	
M-1602	Instrument/Service Air Receiver	3.5	10	V	CS	

TABLE C-2 (Cont)

Heat Exchanger Summary

Item No.	Service	Fluid Shell/Tube	Area (ft^2)	Material Shell/Tube	Comments
Section 200 - Steam Explosion/Wash					
T-201	Vacuum Flash Condenser	H_2O/H_2O (vapor)	2220	CS/SS	Knockback Condenser
Section 300 - Enzyme Production					
T-301A,B	Fermenter No. 1 Recycle Cooler	H_2O/Fermenter Broth	600	CS/SS	Plate and Frame
T-302A,B	Fermenter No. 2 Recycle Cooler	H_2O/Fermenter Broth	400	CS/SS	Plate and Frame
T-303A,B	Air Sparge Cooler	Air/H_2O	1250	CS/CS	
T-304A,B	Air Compressor Intercooler	Air/H_2O			Vendor Package
Section 400 - Hydrolysis					
T-401	Hydrolysis Dilution Cooler	H_2O/H_2O	1000	SS/SS	
Section 500 - Evaporation					
T-501A,B	Evaporator Surface Condenser	H_2O/H_2O	6500	304SS/304SS	
T-502	Evaporator Chest No. 1	Steam/Slurry		304SS/304SS	Vendor Package
T-503	Evaporator Chest No. 2	Steam/Slurry		304SS/304SS	Vendor Package
T-504	Evaporator Chest No. 3	Steam/Slurry		304SS/304SS	Vendor Package
T-505	Evaporator Chest No. 4	Steam/Slurry		304SS/304SS	Vendor Package
T-506	Evaporator Chest No. 5	Steam/Slurry		304SS/304SS	Vendor Package
T-509A-F	Evaporator Feed Heater	Steam/Slurry		304SS/304SS	Vendor Package
Section 600 - Fermentation					
T-601	Fermenter Feed Cooler	H_2O/H_2O-sugar-solids	3055	CS/CS	Two Shells
T-602	Refrigeration Loop Cooler				Vendor Package
T-603	Refrigerant Condenser				Vendor Package
T-604	Fermenter Feed Chiller	H_2O/H_2O-sugar-solids	416	CS/CS	
Section 700 - Distillation					
T-701	Beer Still Feed Preheater	ETOH-H_2O/H_2O	1520	CS/CS	
T-702	Beer Still Trim Condenser	ETOH-H_2O/H_2O	870	CS/CS	
T-703	Recovery Column Overhead Condenser	Entrainer/H_2O	815	CS/CS	
T-704	Ethanol Product Cooler	ETOH/H_2O	520	CS/CS	
T-705	Evaporator Feed Preheater	H_2O/ETOH-H_2O	2700	CS/CS	
T-706	Beer Still Bottoms Cooler	H_2O/ETOH-H_2O	3350	CS/CS	
T-708	Beer Still Reboiler	Steam/ETOH-H_2O	600	CS/CS	
T-709	Recovery Column Reboiler	Steam/H_2O	335	CS/CS	

TABLE C-2 (Cont)

Heat Exchanger Summary

Item No.	Service	Fluid Shell/Tube	Area (ft^2)	Material Shell/Tube	Comments
T-711	Decanter Feed Cooler	Entrainer/H_2O	400	CS/CS	
T-712	Fusel Oil Cooler	H_2O/H_2O	70	CS/CS	
T-713	Anhydrous Column Purge Cooler	Entrainer/H_2O	10	CS/CS	
T-715	Anhydrous Column Overhead Condenser	Steam-ETOH/H_2O	1230	CS/CS	
T-717	Anhydrous Column Hold Tank Feed Cooler	ETOH/H_2O	160	CS/CS	
T-719	Anhydrous Column Reboiler	Steam/ETOH	790	CS/CS	
T-724	Anhydrous Column Hold Tank Vent Condenser	CO_2-ETOH/H_2O	3	CS/CS	
T-726	Beer Still Vent Condenser	CO_2-ETOH/H_2O	55	CS/CS	
T-727	Degasser Drum Vent Condenser	CO_2-ETOH/H_2O	16	CS/CS	
Section 800 - Anaerobic Digestion					
T-801	Digester Feed Cooler	H_2O/Broth			Vendor Package
Section 900 - Boiler					
T-901	Primary Economizer	Flue Gas/H_2O-Steam			Vendor Package
T-902	Superheater	Flue Gas/Steam			Vendor Package
T-903	Air Preheater	Flue Gas/Air			Vendor Package

Tank Summary

Item No.	Service	Material Stored	Size (ft^3)	Comments
Section 900 - Boiler				
Q-901	Lignin Day Bin	Wet Solids	6,300	
Q-902	Ash Silo	Ash	300	
Q-904	No. 2 Fuel Oil Storage Tank	No. 2 Fuel Oil	25,000	
Section 1000 - Wood Feeding				
Q-1001	Truck Receiving Hopper	Wood Chips	5,120	
Q-1002	Surge Bin	Wood Chips	800	
Section 1400 - Product Storage and Unloading				
Q-1401	Denatured Alcohol Storage Tank	ETOH/Gasoline	84,000	
Q-1402	Gasoline Storage Tank	Gasoline	4,423	
Section 1600 - Instrument Air/Fire Protection				
Q-1601	Fire Protection Water Storage	H_2O	53,000	
Q-1602	Hydropneumatic Tank	H_2O	40	Diaphragm Type
Q-1603	Diesel Fuel Oil Tank	Diesel Fuel	67	

TABLE C-3

MOTOR LIST

Item No.	Service	Number Operating	Rated Hp (each)	Total Operating Hp	Comments
Section 100					
P-101A-F	Acid Recycle Pumps	6	75	405	
P-102A-F	Sulfuric Acid Metering Pump	1	0.5	0.5	Intermittent
W-101	Inclined Conveyor Feeder	1	3	2	
W-102	Pretreatment Feed Conveyor	1	3	2	
W-103A,B	Impregnation Vessel Feed Conveyor	2	5	4	
W-104A-F	Impregnation Vessel Screw Conveyor	6	50	40	Intermittent
W-105A-F	Impregnation Product Screw Conveyor	6	10	8	Intermittent
W-106	Central Product Belt Conveyor	1	12.5	10	
Section 200					
P-201A&B	Anaerobic Digester Feed Pump	1	15	11	
P-202A&B	Lignin Centrifuge Feed Pump	1	7.5	7	
P-203A,B	Water/Alkali Wash Feed Pump	2	15	30	
P-204A,B	Water Wash Recycle Pump	2	10	20	
R-201	Vacuum Pump	1	1	1	
W-201	Vibrating Rotary Feeder	1	5	4	
W-202	Washer Screw Conveyor	1	7.5	6	
W-203	Washed Cellulose Lift Conveyor	1	12.5	10	
V-201	Counter-Current Water/Alkali Wash	1	150	120	
Section 300					
G-301A-F	Fermenter Agitators	6	150	720	
G-302	Primary Seed Vessel Agitator	1	3	3	
G-303	Seed Culture Vessel Agitator	1	10	10	
P-301A,B	Fermenter 1 Recycle Pump	2	10	15	
P-302A,B	Fermenter 2 Recycle Pump	2	10	20	
P-303A,B	Fermenter 3 Product Pump	2	10	20	
P-304	Enzyme Seed Pump	1	3	1	
P-305	Seed Culture Pump	1	0.5	0.1	
P-308	Ammonia Pump	1	0.5	0.1	
P-309	CSL Pump	1	0.5	0.1	
R-301A,B	Air Compressor	2	2277	4554	Steam Driven
W-302	Enzyme Fermenter Feed Conveyor	1	30	5	Intermittent

Item No.	Service	Number Operating	Rated Hp (each)	Total Operating Hp	Comments
V-302	Clean In Place System	1	150	50	Intermittent
Section 400					
G-401A-L	Hydrolysis Reactor Agitators	12	100	960	
G-402A,B	Enzyme Recovery Tank Agitators	2	125	200	
G-403A,B	Hydrolysis Recycle Centrifuge	2	500	800	
G-404	Hydrolysis Centrifuge	1	350	280	
P-401A,B	Enzyme Recovery Pump	1	10	9	
P-403A,B&C	Evaporator Feed Pump	2	15	20	
P-404A,B&C	Hydrolysis Recycle Pump	2	5	9	
P-405	Sulfuric Acid Metering Pump	1	0.5	0.5	
P-407A,B	Hydrolysate Pumps	2	30	27	
W-401	Hydrolysis and Enzyme Prod Feed Conveyor	1	40	32	
W-402A,B	Hydrolysis Reactor Feed Conveyor	2	50	40	Intermittent
W-403	Water Wash Conveyor	1	3	2	
W-404	Washer Screw Discharge	1	3	2	
W-405	Lignin Transfer Conveyor	1	4	3	
V-401	Counter-Current Wash	1	300	240	
Section 500					
P-501	Evaporator Feed Pump	1	880	600	
P-502	Evaporator Circulation Pump No. 1	1	880	600	
P-503	Evaporator Circulation Pump No. 2	1	880	600	
P-504	Evaporator Circulation Pump No. 3	1	880	600	
P-505	Evaporator Circulation Pump No. 4	1	880	600	
P-506	Evaporator Circulation Pump No. 5	1	880	600	
P-507	Evaporator Condensate Pump No. 1	1	880	600	
P-508	Evaporator Condensate Pump No. 2	1	880	600	
P-509	Evaporator Condensate Pump No. 3	1	880	600	
R-501	Evaporator Vacuum Pump	1	880	600	

Item No.	Service	Number Operating	Rated Hp (each)	Total Operating Hp	Comments
Section 600					
G-601	Mix Tank Agitator	1			
P-601	Mix Tank Feed Pump	1			
P-602	Immobilized Bead Tank Feed Pump	1			
P-603	Beer Still Feed Tank Feed Pump	1	10	10	
P-604A,B,&C	Refrigerated Water Circ Pump	2	15	4	
P-605A-C,&D	Second Stage Fermenter Feed Pump	3			
R-601	Air Compressor	1	10	5	
R-602	Refrigerant Compressor	1	1233	1233	

*Included as an annual operating cost for the entire fermentation package.

Item No.	Service	Number Operating	Rated Hp (each)	Total Operating Hp	Comments
Section 700					
P-701A&B	Beer Still Reflux Pump	1	7.5	5	
P-702A&B	Anhydrous Column Reflux Pump	1	10	7	
P-703A&B	Recovery Column Reflux Pump	1	3	2	
P-704A&B	Beer Still Reboiler Circulation Pump	1	1.5	1	
P-705A&B	Ethanol Product Pump	1	1.5	1	
P-707A&B	Anhydrous Column Feed Pump	1	5	3	
P-708A&B	Backstillage Pump	1	10	7	
P-709	Anhydrous Column Rerun Pump	1	0.5	0.5	
P-712	Entrainer Makeup Pump	1	0.5	-	Intermittent
P-713A&B	Recovery Column Bottoms Pump	1	0.5	0.5	
P-715	Anhydrous Column Recycle Pump	1	3	2	
P-716	Beer Still Feed Pump	1	20	15	
P-721A&B	Anhydrous Column Reboiler Condensate Pump	1	1.5	1	
P-724	Fusel Oil Product Pump	1	0.5	0.5	
P-725A&B	Recovery Column Feed Pump	1	1.5	1	
Section 800					
P-801A&B	Digester Feed Pump	1	10	8	Vendor Package
P-802	Nutrient Feed Pump	1	0.5	0.5	Vendor Package
P-803	Digested Sludge Pump	1	10	8	
R-801	Methane-Rich Gas Compressor	1	175	140	

Item No.	Service	Number Operating	Rated Hp (each)	Total Operating Hp	Comments
Section 900					
G-904	Lignin Centrifuge	1	300	240	
P-901	Fuel Oil Unloading Pump	1	7.5	0.5	Intermittent
P-902	Fuel Oil Pump	1	2	0.1	Intermittent
R-901	Primary Air Fan	1	1410	1130	
R-902	Induced Draft Fan	1			
W-901	Boiler Screw Feeder	1	3	2	
W-902	Ash Silo Feed Screw Conveyor	1	1.5	1	
W-903	Lignin Day Bin Screw Feeder	1	1	1	
W-904	Boiler Wood Chip Conveyor	1	25	20	
W-905	Boiler Wood Chip Screw Feeder	1	1	1	
Section 1000					
G-1005	Scalping Screen	1	7.5	5	Intermittent
G-1006	Primary Magnetic Separator	1	5	3	Intermittent
G-1007	Secondary Magnetic Separator	1	5	3	Intermittent
G-1008	Stone Trap	1	0.5	0.5	Intermittent
G-1009	Wood Chip Blower	1	60	36	Intermittent
G-1010A,B	Three Deck Chip Screens	2	10	6	Intermittent
G-1012	Storage Silo w/Screw Discharge	1	25	15	Intermittent
W-1001	Unloading Bin Drag Chain	1	75	45	Intermittent
W-1002	Scalping Screen Feed Drag Chain	1	75	45	Intermittent
W-1003	Double Wing Belt Stacker	1	20	12	Intermittent
W-1004	Stacker Feed Conveyor	1	40	24	Intermittent
W-1005A,B	Traveling Scraper/Reclaimer	2	130	78	Intermittent
W-1006	Storage Pile Transfer Conveyor	1	25	15	Intermittent
W-1007A,B	Reclaimer Belt Conveyor	2	30	18	Intermittent
W-1008	Wood Chip Elevating Conveyor	1	25	15	Intermittent
W-1009	Fines Transfer Conveyor	1	2	1	Intermittent
W-1010	Large Chip Transfer Conveyor	1	2	1	Intermittent
W-1011	Medium Chip Transfer Conveyor	1	3	2	Intermittent
W-1012	Clean Wood Chip Transfer Conveyor	1	25	15	Intermittent
W-1013	Boiler Fuel Transfer Conveyor	1	20	16	
Section 1100					
G-1102A-F	Induced Draft Fan	6	60	360	
P-1101A,B&C	Cooling Water Circulation Pump	2	250	468	

Item No.	Service	Number Operating	Rated Hp (each)	Total Operating Hp	Comments
V-1101	Inhibitor Feed System	1	2.5	2.5	
V-1102	Acid Feed System	1	2.5	2.5	
Section 1200					
G-1204A,B,&C	Belt Filter Press	2	5	8	
G-1209	Neutralization Tank Agitator	1	0.5	0.1	Intermittent
G-1210	Primary Clarifier Rake	1	0.5	0.5	
G-1211	Secondary Clarifier Rake	1	0.5	0.5	
G-1212A-H	Aeration Basin Agitator	8	60	384	
G-1213	Final Clarifier Rake	1	0.5	0.5	
G-1214	Sludge Thickener Rake	1	0.5	0.5	
G-1215	Polymer Feed Tank Agitator	1	1	1	
G-1216	Sludge Mixing Tank Agitator	1	2	2	
G-1217A,B,&C	Trickling Filter Fans	2	1	2	
P-1201A&B	Settling Basin Effluent Pump	1	3	1	Intermittent
P-1202A&B	Separator Water Pump	1	3	1	Intermittent
P-1203A&B	Sulfuric Acid Feed Pump	1	1	0.5	Intermittent
P-1204A&B	Sodium Hydroxide Feed Pump	1	1	0.5	Intermittent
P-1205A&B	Primary Clarifier Sludge Pump	1	5	4	Intermittent
P-1206	Polymer Transfer Pump	1	1	0.5	Intermittent
P-1207A&B	Polymer Feed Pump	1	1	0.5	Intermittent
P-1208A&B	Filtrate Transfer Pump	1	5	4	
P-1209A,B,&C	Trickling Filter Feed Pump	2	70	98	
P-1210A&B	Secondary Clarifier Sludge Pump	1	5	4	
P-1211A&B	Final Clarifier Sludge Pump	1	5	4	
P-1212A&B	Neutralization Feed Pump	1	2	1	Intermittent
P-1213A&B	Neutralized Effluent Pumps	1	3	1	Intermittent
P-1214A&B	Ammonium Hydroxide Feed Pump	1	1	1	
P-1215A&B	Phosphoric Acid Feed Pump	1	1	1	
P-1216A&B	Clean Water Discharge Pump	1	25	18	
P-1217A&B	Belt Filter Washwater Pump	1	3	1	Intermittent
P-1218	CO_2 Wash Column Pump	1	0.75	0.5	
P-1219	Vent Scrubber Pump	1	0.5	0.5	
P-1220A&B	Sludge Thickener Underflow Pump	1	3	2	
P-1221A,B,&C	Sludge Transfer Pump	2	2	3	
R-1201	CO_2 Wash Column Blower	1	150	120	
R-1201	Vent Scrubber Blower	1	15	12	
W-1201	Thickened Sludge Conveyor	1	2	1	Intermittent

TABLE C-3 (Cont)

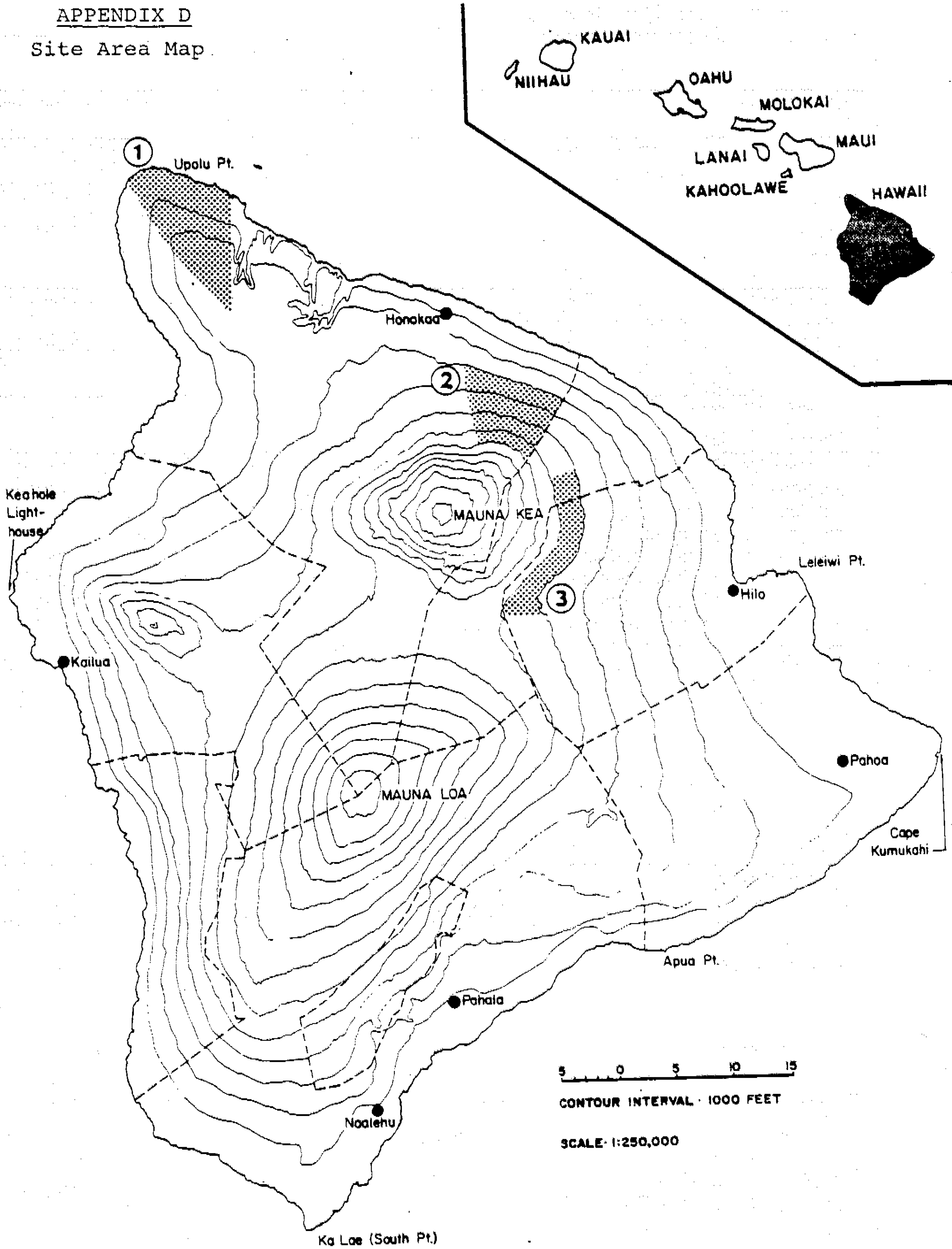
Item No.	Service	Number Operating	Rated Hp (each)	Total Operating Hp	Comments
Section 1300					
P-1301	Sodium Hydroxide Feed Pump	1	1	0.1	Intermittent
P-1302	Sulfuric Acid Feed Pump	1	0.5	0.1	Intermittent
Section 1400					
P-1401A,B	Denatured Alcohol Loading Pump	2	15	2	Intermittent
P-1402	Gasoline Unloading Pump	1	10	0.5	Intermittent
P-1403A,B	Gasoline Metering Pump	1	0.5	0.1	
Section 1500					
P-1501	Caustic Pump	1	0.5	-	Intermittent
P-1502	Hydrochloric Acid Pump	1	0.5	-	Intermittent
P-1503	Deaerator Feed Pump	1	5.0	4	
P-1504	Boiler Feedwater Pump	1	250	206	
Section 1600					
P-1601	Motor Driven Fire Pump	-	300	-	Not normally operating
P-1602	Jockey Fire Pump	1	10	8	
P-1604A,B	Diesel Fuel Oil Pump	1	5	-	Not normally operating
R-1601	Instrument Air Compressor	1	125	100	
R-1602	Instrument Service Air Compressor	1	125	100	

APPENDIX D

MAP OF THE ISLAND OF HAWAII

APPENDIX D

Site Area Map



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15. Supplementary Notes

Technical Monitor: John Wright

16. Abstract (Limit: 200 words) This report presents an economic feasibility study of an enzyme-based ethanol plant. The objectives of the study were to determine the current economic status of the conversion of lignocellulose to ethanol via enzymatic hydrolysis and to make recommendations for further R&D. The results include an integrated process design, a capital cost estimate, an investment analysis, and R&D recommendations. The site for the enzyme-based ethanol plant is on the island of Hawaii, near the city of Hilo. The fill-scale plant will be capable of producing 15 million gallons per year of fuel-grade ethanol from eucalyptus wood chips.

17. Document Analysis

a. Descriptors Hydrolysis, enzymatic hydrolysis, cellulolytic activity, ethanol, ethanol fuels, biomass conversion plants, eucalyptuses, feasibility studies, economic analysis

b. Identifiers/Open-Ended Terms

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